

# Chapter 3

## THERMAL and CATALYTIC INCINERATORS

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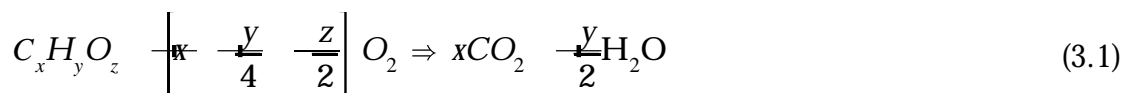
## 3.1 Introduction

Incineration, like carbon adsorption, is one of the best known methods of industrial gas waste disposal. Unlike carbon adsorption, however, incineration is an ultimate disposal method in that the objectionable combustible compounds in the waste gas are converted rather than collected. On the other hand, carbon adsorption allows recovery, of organic compounds which may have more value as chemicals than just their heating value. A major advantage of incineration is that virtually any gaseous organic stream can be incinerated safely and cleanly, provided proper engineering design is used.

The particular application of both thermal and catalytic incineration to gaseous waste streams containing volatile organic compounds (VOCs) is discussed here. The U.S. Environmental Protection Agency defines any organic compound to be a VOC unless it is specifically determined *not* to be a VOC. Indeed, a number of organics (e.g., methane) are specified as not being VOCs. Although both VOC and non-VOC organic compounds are combustible and are therefore important in the design of the incinerator, this distinction is important since it is the control of VOCs that is regulated.

## 3.2 Process Description

Seldom is the waste stream to be combusted a single organic compound. Rather, it is common to have a complex mixture of organic compounds. This mixture is typically analyzed for carbon, hydrogen, oxygen, and other elements; and an empirical formula is developed which represents the mixture. Combustion of such a mixture of organic compounds containing carbon, hydrogen, and oxygen is described by the overall exothermic reaction:



The complete combustion products  $CO_2$  and  $H_2O$  are relatively innocuous, making incineration an attractive waste disposal method. When chlorinated sulfur-containing compounds are present in the mixture, the products of complete combustion include the acid components  $HCl$  or  $SO_2$ , respectively, in addition to  $H_2O$  and  $CO_2$ . In general, these streams would require removal of the acid components by a scrubber unit, which could greatly affect the cost of the incineration system. (The sizing and costing of these scrubbers is covered in the "Wet Scrubbers" chapter of this *Manual*.)

The heart of an incinerator system is a combustion chamber in which the VOC-containing waste stream is burned. Since the inlet waste gas stream temperature is generally much lower than that required for combustion, energy must be supplied to the incinerator to raise the waste gas temperature. Seldom, however, is the energy released by the combustion of the total organics (VOCs and others) in the waste gas stream sufficient to raise its own temperature to the desired levels, so that auxiliary fuel (e.g., natural gas) must be added.

The combustion of the waste gases may be accomplished in a thermal incinerator or in a catalytic incinerator. In the catalytic incinerator a catalyst is used to increase the rate of the combustion reaction, allowing the combustion to occur at lower temperatures. Because the catalytic process operates at a lower temperature than the thermal process, less auxiliary fuel may be required in the catalytic process to preheat the waste gas.

Auxiliary fuel requirements may also be decreased, and energy efficiency improved, by providing heat exchange between selected inlet streams and the effluent stream. The effluent stream containing the products of combustion, along with any inerts that may have been present in or added to the inlet streams, can be used to preheat the incoming waste stream, auxiliary air, or both via a "primary", or recuperative, heat exchanger. It is useful to define the fractional energy recovery by the preheater, or primary, heat exchanger as follows:

$$\text{Fractional energy recovery} = \frac{\text{Energy actually recovered flue gas}}{\text{Maximum energy recoverable if flue gas approached lowest temperature available to heat exchanger}} \quad 3.2$$

The energy actually recovered, the numerator of Equation 3.2, is the increase in sensible heat of the gas, *i.e.*, waste gas or waste gas plus dilution air, being heated. The maximum energy recoverable would be the decrease in sensible heat of the flue gas, if it were cooled to the temperature of the incoming waste gas. While this maximum energy recovery would be attained only with a very large heat exchanger, the concept of fractional energy recovery is useful in expressing the extent of the improvement in energy efficiency using a "primary" heat exchanger.

Energy efficiency can be further improved by placing another ("secondary") exchanger downstream of the primary exchanger to recover additional energy from the effluent stream (*e.g.*, to generate low pressure process steam or hot water). However, secondary energy recovery is generally not used, unless there is a specific on site use for the steam or hot water.

The majority of industrial gases that contain VOCs are dilute mixtures of combustible gases in air. In some applications, such as air oxidation processes, the waste gas stream is very deficient in oxygen. Depending on the oxygen content of the waste stream, auxiliary air may be required to combust the total organic content of the waste gas as well as any auxiliary fuel that has been used.

The concentration of combustible gas in the waste gas stream plays an integral role in the design and operation of an incinerator. From a cost standpoint, the amount of air in excess of the stoichiometric amounts should be minimized. For safety reasons, however, any mixture within the flammability limits, on either the fuel-rich or fuel-lean side of the stoichiometric mixture, presents an unacceptable fire hazard as a feed stream to the incinerator. The lower, or fuel-lean, explosive limit (LEL) of a given organic compound defines the minimum concentration of that compound in air that can produce more energy than is needed to raise its own temperature to the ignition point (*i.e.*, ignite). Similarly, the upper, or fuel-rich, explosive

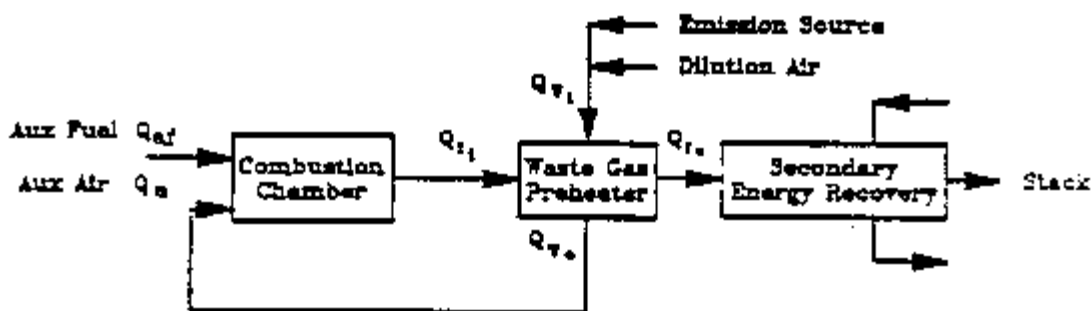
limit (UEL) represents the highest concentration of the organic in air that is ignitable. In the latter case, air is limiting the reaction. Both the LEL and the UEL are measured at ambient conditions. Empirically, it has been found that mixtures of hydrocarbons in air at their LEL have a heating value of approximately 50 Btu/scf.

Since the majority of industrial waste gases that contain VOCs are dilute mixtures of combustible gases in air, their heating value is low and their oxygen content exceeds that required to combust both the waste organics (VOCs and others) and the auxiliary fuel. If a waste gas above 50 percent LEL (about 25 Btu/scf) is encountered, it must be diluted to satisfy fire insurance regulations. Generally, the streams are brought to below 25 percent LEL, although concentrations from 25 percent to 50 percent are permitted provided the waste stream is continuously monitored by LEL monitors. Because air is the usual diluent gas, care must be taken with preheating the diluted stream so that it remains below about 1200°F. (See discussion below on preheating.) A table showing LEL, UEL, and heats of combustion for selected organic compounds is given in Appendix 3A.

The goal of any incineration system is to control the amount of VOCs released to the environment. Performance of a control device such as an incinerator can be described by a control efficiency, defined according to the following equation:

$$\text{Control Eff., \%} = \left[ \frac{\text{Inlet mass rate VOC} - \text{Outlet mass rate VOC}}{\text{Inlet mass rate VOC}} \right] \times 100 \quad 3.3$$

It is important to note, however, that incomplete combustion of the inlet VOCs could result in the formation of other VOCs not originally present. For example, the incomplete oxidation of dichloroethane can yield vinyl chloride. Both of these compounds are VOCs. The definition given in Equation 3.3 would still be meaningful, however, as long as the newly formed VOC (e.g., vinyl chloride) is detected. This situation necessitates the complete chemical analysis of the inlet and outlet gas streams to confirm compliance with State and Federal regulations.



**Figure 3.1:** Thermal Incinerator - General Case

Performance of an incinerator can also be measured solely by the *outlet* VOC concentration, usually in ppmv.

There are a number of different incinerator designs. These designs can be broadly classified as thermal systems and catalytic systems. Thermal systems may be direct flame incinerators with no energy recovery, flame incinerators with a recuperative heat exchanger, or regenerative systems which operate in a cyclic mode to achieve high energy recovery. Catalytic systems include fixed-bed (packed-bed or monolith) systems and fluid-bed systems, both of which provide for energy recovery. The following sections discuss design aspects of these systems.

### 3.2.1 Thermal Incinerators

The heart of the thermal incinerator is a nozzle-stabilized flame maintained by a combination of auxiliary fuel, waste gas compounds and supplemental air added when necessary (see Figure 3.1). Upon passing through the flame, the waste gas is heated from its inlet temperature (e.g., 100°F) to its ignition temperature. The ignition temperature varies for different compounds and is usually determined empirically. It is the temperature at which the combustion reaction rate (and consequently the energy production rate) exceeds the rate of heat losses, thereby raising the temperature of the gases to some higher value. Thus, any organic/air mixture will ignite if its temperature is raised to a sufficiently high level.

The organic-containing mixture ignites at some temperature between the preheat temperature and the reaction temperature. That is, ignition, as defined in this section, occurs at some point during the heating of a waste stream as it passes through the nozzle-stabilized flame regardless of its concentration. The mixture continues to react as it flows through the combustion chamber.

The required level of VOC control of the waste gas that must be achieved within the time that it spends in the thermal combustion chamber dictates the reactor temperature. The shorter the residence time, the higher the reactor temperature must be. The nominal residence time of the reacting waste gas in the combustion chamber is defined as the combustion chamber volume divided by the volumetric flow rate of the gas. Most thermal units are designed to provide no more than 1 second of residence time to the waste gas with typical temperatures of 1,200 to 2,000°F. Once the unit is designed and built, the residence time is not easily changed, so that the required reaction temperature becomes a function of the particular gaseous species and the desired level of control. Table 3.1 illustrates the variability in (theoretical) reactor temperatures that is required to destroy 99.99 percent of the inlet mass of various noxious compounds with excess air for a 1-second reactor residence time [1].

**Table 3.1:** Theoretical Reactor Temperatures Required for 99.99 Percent Destruction by Thermal Incineration for a 1-Second Residence Time\*

Compound	Temperature, °F
acrylonitrile	1,344
allyl chloride	1,276
benzene	1,350
chlorobenzene	1,407
1,2-dichloroethane	1,368
methyl chloride	1,596
toluene	1,341
vinyl chloride	1,369

\*Reference [1]

These temperatures cannot be calculated a priori, although incinerator vendors can provide guidelines based on their extensive experience. In practice, most streams are mixtures of compounds, thereby further complicating the prediction of this temperature. Other studies [2,3,4], which are based on actual field test data, show that commercial incinerators should generally be run at 1600°F with a nominal residence time of 0.75 seconds to ensure 98% destruction of non-halogenated organics. In some States the reactor temperature and residence time of the unit are specified rather than attempting to measure actual levels of VOC control. The selected temperature must be maintained for the full, selected residence time for combustion to be complete.

These three studies also conclude that *mixing* is a critical factor in determining the destruction efficiency. Even though it cannot be measured, mixing is a factor of equal or even greater importance than other parameters, such as temperature. The most feasible and efficient way to improve the mixing in an incinerator is to adjust it after start-up. The 98% control level discussed in the previous paragraph presumes such an adjustment.



Ultimately, once the unit is built, it is the responsibility of the user to operate and maintain the incinerator to insure compliance with applicable regulations.

### **3.2.1.1 Direct Flame Incinerators**

Many configurations of thermal incinerators exist with the same goal—to raise the VOC-containing stream to the desired reaction temperature and hold it there for the given reaction time to achieve the required destruction efficiency. The simplest example of such a system is the direct flame incinerator. With reference to Figure 3.1, the direct flame incinerator is comprised only of the combustion chamber. The waste gas preheater and the secondary energy recovery heat exchanger are energy recovery devices and are not included as part of the direct flame incinerator.

### **3.2.1.2 Recuperative Incinerators**

Recuperative incinerators have improved energy efficiency as a result of placing heat exchangers in the hot outlet gas streams. With reference to Figure 3.1, the recuperative incinerator is comprised of the combustion chamber, the waste gas preheater, and, if appropriate, the secondary, energy recovery heat exchanger.

**Primary Energy Recovery (Preheating Inlet Streams)** Considerable fuel savings can be realized by using the exit (product) gas to preheat the incoming feed stream, combustion air, or both via a heat exchanger, as shown in Figure 3.1 in the so-called "recuperative" incinerator. These heat exchangers can recover up to 70% of the energy (enthalpy) in the product gas.

The two types of heat exchangers most commonly used are *plate-to-plate* and *shell-and-tube*. Plate-to-plate exchangers offer high efficiency energy recovery at lower cost than shell-and-tube designs. Also, because of their modular configuration, plate-to-plate units can be built to achieve a variety of efficiencies. But when gas temperatures exceed 1000°F, shell-and-tube exchangers usually have lower purchase costs than plate-to-plate designs. Moreover, shell-and-tube exchangers offer better long-term structural reliability than plate-to-plate units. [5] In any case, because most incinerators installed are packaged units, the design (and cost) of the recuperative heat exchangers have already been incorporated.

Most heat exchangers are not designed to withstand high temperatures, so that most of the energy needed to reach ignition is supplied by the combustion of fuel in the combustion chamber and only moderate preheat temperatures are sought in practice (< 1200°F).

**Secondary Energy Recovery (Additional Waste Energy Recovery)** It should be noted, however, that at least some of the energy added by auxiliary fuel in the traditional thermal units (but not recovered in preheating the feed stream) can still be recovered. Additional heat exchangers can be added to provide process heat in the form of low pressure steam or hot water for on-site application. Obviously, an in-plant use for such low level energy is needed to realize these savings.

The need for this higher level of energy recovery will be dependent upon the plant site. The additional heat exchanger is often provided by the incineration unit vendor. The cost of this

additional heat exchanger may be estimated via standard heat exchanger correlations and should be added to the costs estimated using the cost correlations in this chapter.

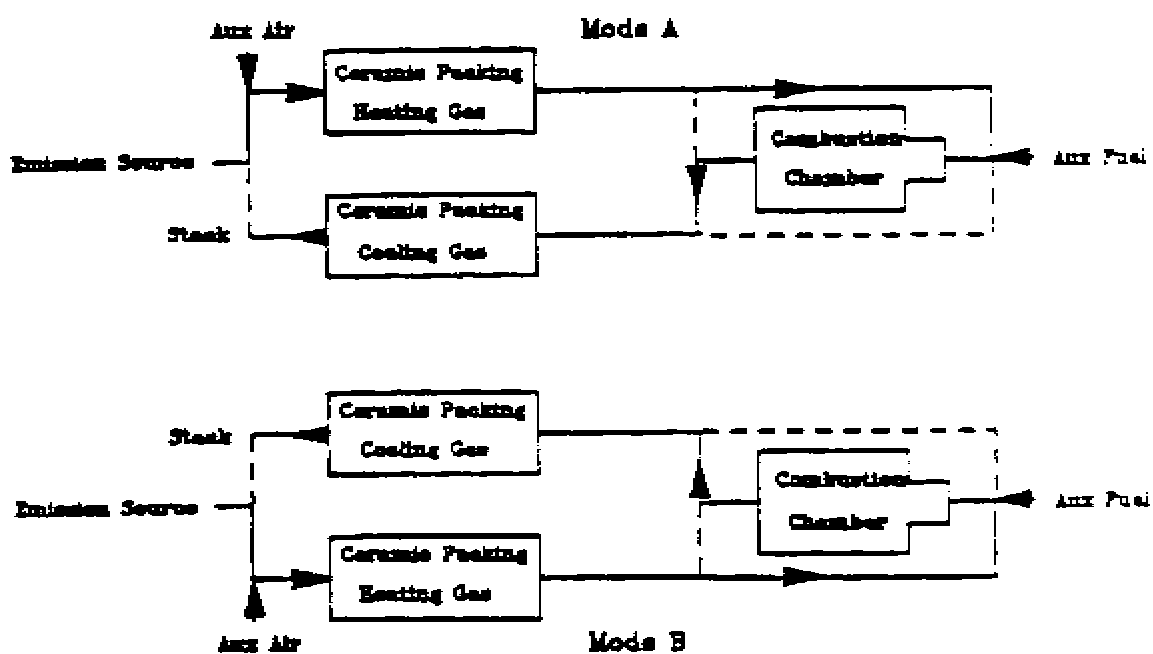
### **3.2.1.3 Regenerative Incinerators**

A distinction in thermal incinerators can now be made on the basis of this limitation in the preheat temperature. The traditional approach to energy recovery in the units (shown schematically in Figure 3.1) still requires a significant amount of auxiliary fuel to be burned in the combustion chamber when the waste gas heating values are too low to sustain the desired reaction temperature at the moderate preheat temperature employed. Additional savings can, under these conditions, be realized in units with more complete transfer of exit-stream energy. This is the concept behind the so-called excess-enthalpy or regenerable burner systems. These systems use direct contact heat exchangers constructed of a ceramic material that can tolerate the high temperatures needed to achieve ignition of the waste stream.

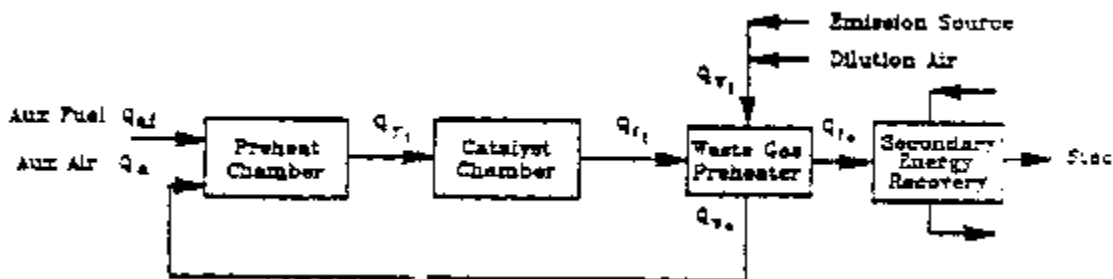
The operation of the regenerative system is illustrated in Figure 3.2. The inlet gas first passes through a hot ceramic bed thereby heating the stream (and cooling the bed) to its ignition temperature. If the desired temperature is not attainable, a small amount of auxiliary fuel is added in the combustion chamber. The hot gases then react (releasing energy) in the combustion chamber and while passing through another ceramic bed, thereby heating it to the combustion chamber outlet temperature. The process flows are then switched, now feeding the inlet stream to the hot bed. This cyclic process affords very high energy recovery (up to 95%).

The higher capital costs associated with these high-performance heat exchangers and combustion chambers may be offset by the increased auxiliary fuel savings to make such a system economical. The costs of these regenerative units will be given separately in the cost correlations presented in Section 3.5. Regenerative incinerators are not packaged units but are field-erected only. Accordingly, the costs given in Section 3.5 for regenerative units are for field-erected units.





**Figure 3.2:** Regenerable-type Thermal Incinerator



**Figure 3.3:** Catalytic Incinerator

### 3.2.2 Catalytic Incinerators

Catalytic incinerators employ a bed of active material (catalyst) that facilitates the overall combustion reaction given in Equation 3.1. The catalyst has the effect of increasing the reaction rate, enabling conversion at lower reaction temperatures than in thermal incinerator units. Nevertheless, the waste stream must be preheated to a temperature sufficiently high (usually from 300 to 900°F) to initiate the oxidation reactions. The waste stream is preheated either directly in a preheater combustion chamber or indirectly by heat exchange with the incinerator's effluent or other process heat or both (Figure 3.3). The preheated gas stream is then passed over the catalyst bed. The chemical reaction (combustion) between the oxygen in the gas stream and the gaseous pollutants takes place at the catalyst surface. Catalytic incineration can, in principle, be used to destroy essentially any oxidizable compound in an air stream. However, there are practical limits to the types of compounds that can be oxidized due to the poisoning effect some species have on the catalyst. These limits are described below. In addition, most configurations require a low heating value of the inlet gas and a particulate content which is less than some small value.

Until recently, the use of catalytic oxidation for control of gaseous pollutants has really been restricted to organic compounds containing only carbon, hydrogen and oxygen. Gases containing compounds with chlorine, sulfur, and other atoms that may deactivate the supported noble metal catalysts often used for VOC control were not suitably controlled by catalytic oxidation systems. Catalysts now exist, however, that are tolerant of such compounds. Most of these catalysts are single or mixed metal oxides, often supported by a mechanically strong carrier such as  $\gamma$ -alumina. Perhaps most of the development of poison-tolerant catalysts has focused on the oxidation of chlorine-containing VOCs. These compounds are widely used as solvents and degreasers and are often the subject of concern in VOC control. Catalysts such as chromia/alumina [6, 7], cobalt oxide [8], and copper oxide/manganese oxide [9] have been used for oxidation of gases containing chlorinated compounds. Platinum-based catalysts are active for oxidation of sulfur containing VOCs, although they are rapidly deactivated by the presence of chlorine. Compounds containing atoms such as lead, arsenic, and phosphorous should, in general, be considered poisons for most oxidation catalysts. Nevertheless, their concentration may be sufficiently low so that the rate of deactivation and therefore, the catalyst replacement costs, could be low enough to consider catalytic oxidation.

**Table 3.2:** Catalyst Temperatures Required for Oxidizing 80% of Inlet VOC to CO<sub>2</sub>, °F for Two Catalysts

Compound	Temperature, °F	
	CO <sub>3</sub> O <sub>4</sub>	Pt - Honeycomb
acrolein	382	294
n-butanol	413	440
n-propylamine	460	489
toluene	476	373
n-butyric acid	517	451
1,1,1-trichloroethane	661	> 661
dimethyl sulfide	—	512

As was the case for thermal units, it is impossible to predict *a priori* the temperature and residence time (*i.e.*, inverse space velocity) needed to obtain a given level of conversion of a VOC mixture in a catalytic oxidation system. For example, Table 3.2 from Pope *et al.* [8] shows the temperature needed for 80% conversion of a number of VOCs over two oxidation catalysts in a specific reactor design. This table shows that the temperature required for this level of conversion of different VOCs on a given catalyst and of the same VOC on different catalysts can vary significantly.

Particulate matter, including dissolved minerals in aerosols, can rapidly blind the pores of catalysts and deactivate them over time. Because essentially all the active surface of the catalyst is contained in relatively small pores, the particulate matter need not be large to blind the catalyst. No general guidelines exist as to particulate concentration and particulate size that can be tolerated by catalysts because the pore size and volume of catalysts vary greatly.

The volumetric gas flow rate and the concentration of combustibles in the gas flowing to the catalytic incinerator should be constant for optimal operation. Large fluctuations in the flow rate will cause the conversion of the VOCs to fluctuate also. Changes in the concentration or type of organics in the gas stream can also affect the overall conversion of the VOC contaminants. These changes in flow rate, organics concentration, and chemical composition are generally the result of upsets in the manufacturing process generating the waste stream. It may be uneconomical to change the process for the sake of making the operation of the catalytic incinerator feasible. In such cases, thermal incinerators (discussed earlier in this chapter) or carbon adsorption (discussed in Chapter 4 of this *Manual*) should be evaluated as alternative control technology.

The method of contacting the VOC-containing stream with the catalyst serves to distinguish catalytic incineration systems. Both fixed-bed and fluid-bed systems are used.

### **3.2.2.1 Fixed-Bed Catalytic Incinerators**

Fixed-bed catalytic incinerators may use a monolith catalyst or a packed-bed catalyst. Each of these is discussed below.

**Monolith Catalyst Incinerators** The most widespread method of contacting the VOC-containing stream with the catalyst is the catalyst monolith. In this scheme the catalyst is a porous solid block containing parallel, non-intersecting channels aligned in the direction of the gas flow. Monoliths offer the advantages of minimal attrition due to thermal expansion/contraction during startup/shutdown and low overall pressure drop.

**Packed-Bed Catalytic Incinerators** A second contacting scheme is a simple packed-bed in which catalyst particles are supported either in a tube or in shallow trays through which the gases pass. The first scheme is not in widespread use due to its inherently high pressure drop, compared to a monolith, and the breaking of catalyst particles due to thermal expansion when the confined catalyst bed is heated/cooled during startup/shutdown. However, the tray type arrangement, where the catalyst is pelletized is used by several industries (e.g., heat-set web-offset printing). Pelletized catalyst is advantageous where large amounts of such contaminants as phosphorous or silicon compounds are present. [10]

### **3.2.2.2 Fluid-Bed Catalytic Incinerators**

A third contacting pattern between the gas and catalyst is a fluid-bed. Fluid-beds have the advantage of very high mass transfer rates, although the overall pressure drop is somewhat higher than for a monolith. An additional advantage of fluid-beds is a high bed-side heat transfer as compared to a normal gas heat transfer coefficient. This higher heat transfer rate to heat transfer tubes immersed in the bed allows higher heat release rates per unit volume of gas processed and therefore may allow waste gas with higher heating values to be processed without exceeding maximum permissible temperatures in the catalyst bed. In these reactors the gas phase temperature rise from gas inlet to gas outlet is low, depending on the extent of heat transfer through imbedded heat transfer surfaces. The catalyst temperatures depend on the rate of reaction occurring at the catalyst surface and the rate of heat exchange between the catalyst and imbedded heat transfer surfaces.

As a general rule, fluid-bed systems are more tolerant of particulates in the gas stream than either fixed-bed or monolithic catalysts. This is due to the constant abrasion of the fluidized catalyst pellets, which helps remove these particulates from the exterior of the catalysts in a continuous manner.

A disadvantage of a fluid-bed is the gradual loss of catalyst by attrition. Attrition-resistant catalysts have been developed to overcome this disadvantage. [11]

### **3.2.3 Other Considerations: Packaged versus Field-Erected Units, Auxiliary Equipment**

#### **3.2.3.1 Packaged vs. Field-Erected Units**

With the exception of regenerative incinerators, the equipment cost correlations included in this chapter are for packaged units only. They are *not* valid for field-erected units. For regenerative incinerators, the correlations are valid for field-erected units only. Packaged units are units that have been shop fabricated and contain all elements necessary for operation, except for connection to facilities at the site, e.g., utilities. The elements include the combustion chamber, preheater, instrumentation, fan, and the necessary structural steel, piping, and electrical equipment. This equipment is assembled and mounted on a "skid" to facilitate installation on a foundation at the plant site. Tie-in to the local emission source is not part of the packaged unit. Units are usually sized to handle flow rates of <20,000 scfm, but can be built to accommodate flow rates up to 50,000 scfm. The cost correlations in this chapter are valid to 50,000 scfm for packaged units, except for fluid-bed units which are valid to 25,000 scfm.

Conversely, field-erected units may be built to any desired size. The combustion chamber, preheater, and other equipment items are designed and fabricated individually, and assembled at the site. However, both the equipment and installation costs of field-erected units are typically higher than those for equivalent-sized packaged units because the factors that improve efficiency of shop-fabrication, such as uniform working environment, availability of tools and equipment, and more efficient work scheduling, are generally not available in the field.

#### **3.2.3.2 Acid Gas Scrubbers**

The final outlet stream of any incineration system may contain certain pollutants that must be removed. The combustion of sulfur-containing compounds results in  $\text{SO}_2$ , while chlorinated compounds yield  $\text{Cl}_2$  and  $\text{HCl}$  in the product stream. These acid gases must be removed from the gas stream if they are present at significant concentrations (regulations for limits on these gases vary from state to state). This removal can be effected in, for instance, a packed-bed gas absorber (vertical scrubber) in which the flue gas is contacted with a caustic scrubbing liquid. For fluid-bed catalytic reactors, venturi scrubbers are often used because they provide for particulate removal as well as acid gas scrubbing. In most cases adding a scrubber or absorber significantly increases the cost of the incineration unit, sometimes by a factor of two. Costing of absorbers is discussed in the "Gas Absorbers" chapter (Chapter 9) of this *Manual*.

If chlorinated VOCs are present in the waste gas, heat exchangers may require special materials of construction. This added expense is not included in the costing procedures outlined in this chapter.

#### **3.2.3.3 Heat Exchangers (Preheaters and Other Waste Energy Recovery Units)**

For the thermal and catalytic units having some degree of energy recovery, the cost of the primary heat exchanger is included in the cost, and its design is usually done by the incineration unit vendor. The cost correlations presented in this chapter include units both with and without



energy recovery. Secondary energy recovery, if desired, requires an additional heat exchanger, which is also often provided by the incineration unit vendor. Costing procedures for secondary energy recovery are not included in this chapter.

#### **3.2.3.4 Other Auxiliary Equipment**

Additional auxiliary equipment such as hoods, ductwork, precoolers, cyclones, fans, motors, and stacks are addressed separately in other chapters of this *Manual*.

### **3.2.4 Technology Comparison**

Both the thermal and catalytic incineration systems are designed to provide VOC control through combustion at a level in compliance with applicable state and federal requirements. Given the wide range of options available, however, it is obvious that not every incinerator will fulfill these requirements at the same cost. This section presents a first step toward deciding how best to deal with VOC emission abatement using incinerators considering some qualitative factors pertinent to the types of incinerators described in this chapter. It is the intent of the remainder of Chapter 3 to provide a method by which the cost of VOC control for a particular application can be calculated.

[illegible]

- Direct Flame Incinerator
- Recuperative Incinerator (Direct Flame with Recuperative Heat Exchanger)
- Regenerative Incinerator Operating in a Cyclic Mode

- Fixed-Bed
  - Monolith
  - Packed-Bed
- Fluid-Bed

- The fouling of the catalyst in a catalytic system is a possibility. Poisons to the system include heavy metals, phosphorous, sulfur and most halogens, although catalysts have been developed that are chlorine resistant.
- The possibility of process upsets that could release any of the above poisons or cause fluctuations in the heating value to the incinerator would favor a thermal system.
- Except for the No.2 grade, fuel oil should not be considered as auxiliary fuel to a catalytic system due to the sulfur and vanadium it may contain. [10]

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- Thermal units generally require more auxiliary fuel than catalytic units and operate at temperatures that are roughly 1000°F higher. This difference in fuel requirement increases as the heating value of the waste stream decreases.

In general, a trade-off exists between the higher capital costs of catalytic incinerators and the higher operating costs of thermal incinerators. This difference will be illustrated by a design example presented in Section 3.4 which treats both technologies.

### 3.3 General Treatment of Material and Energy Balances

In the sizing and costing of the incinerator and the calculation of the auxiliary fuel requirements, it is necessary to make material and energy balances around the entire incinerator unit and around certain parts of the unit, such as the combustion chamber or the preheater. This section presents a general approach to making these balances.

These balances are based on the law of conservation of mass and energy. They can be stated in general equation form as

$$In - Out + Generation = Accumulation \quad (3.4)$$

Because the incineration process is a steady-state process, the accumulation term is zero and the equation becomes

$$In - Out + Generation = 0$$

For *mass* balances it is useful to restrict the balances to be made on the mass of each atomic species so that for mass balances the generation term becomes zero. However, because the combustion reaction liberates energy, the *energy* balances around equipment where combustion takes place would include a generation term. Thus, the simplified equations are

$$In - Out = 0, \text{ for steady-state mass balances} \quad (3.5)$$

$$In - Out + Generation = 0, \text{ for steady-state energy balances} \quad (3.6)$$

For the incineration process the two terms *In* and *Out* are generally mass terms (for a mass balance) of the form,

$$\rho Q$$

where

$$\begin{aligned} \rho &= \text{density (mass per unit volume)} \\ Q &= \text{volumetric flow rate (volume per unit time)} \end{aligned}$$

or sensible heat terms (for an energy balance) of the form,

$$\rho Q C_p (T - T_{ref})$$

where

$$\begin{array}{ll} C_p & = \text{heat capacity} \\ T & = \text{temperature} \end{array}$$

The reference temperature,  $T_{ref}$ , is often taken to be zero or the temperature of a convenient stream, e.g., the inlet gas stream, in whatever units  $T$  is in, so the  $T_{ref}$  term may not appear in the equations. When the reference temperature is taken as zero, the sensible heat terms become

$$\rho Q C_p T.$$

Energy losses,  $H_L$ , are also part of the *Out* term and, for the incinerator process, are taken here to be 10% of the total energy *input* to the incinerator.

For the incineration process, the generation term for energy balances accounts for the energy released through the combustion reactions. This term is generally of the form

$$\rho Q (-\Delta h_c)$$

where

$$(-\Delta h_c) = \text{heat of combustion.}$$

### 3.4 Design Procedures

The following procedure is designed to provide parameters for use in developing a study cost estimate (accuracy  $\pm 30\%$ ). The principal parameters of interest are

- flue gas flow rate, upon which all the equipment cost correlations are based.
- auxiliary fuel requirement, which is important in estimating annual operating costs.

For applications which involve control of waste gas streams that are *dilute mixtures of VOCs in air* (> 20% oxygen in the waste gas stream), the flue gas flow rate is greater than the inlet waste gas flow rate by the amount of auxiliary fuel and the increase in the moles of gas as a result of the combustion reaction. Because these two factors usually cause only small increases in flow rate, a number of simplifying assumptions can be made in the design calculations. For

applications where diluent air must be used to adjust the combustible concentration in the waste gas to 25% LEL and where auxiliary fuel and auxiliary combustion air are needed, more complete mass and energy balances must be made.

The design procedure illustrated below is for waste gas streams that are dilute mixtures of VOCs in air (> 20% oxygen in the waste gas stream). In this discussion the design procedure will be illustrated by a sample problem that will be solved step-by-step.

### 3.4.1 Steps Common to Thermal and Catalytic Units

**Step 1 - Establish design specifications** The first step in the design procedure is to determine the specifications of the incinerator and the waste gas to be processed. The following parameters of the waste gas stream at the emission source must be available:

- Volumetric flow rate, scfm—Standard conditions are normally 77°F and 1 atm. pressure
- Temperature
- Oxygen content
- Chemical composition of the combustibles
- Inerts content
- Heating value—In cases the heating value may act as a surrogate for the chemical composition of the combustibles. This is particularly true for dilute mixtures of combustibles in air.
- Particulate content—The particulate content is important if catalytic incinerators are to be coated. An upstream filter may suffice if particulate content is too high. Fluid-bed catalytic incinerators can tolerate higher particulate contents than fixed-bed catalytic incinerators.

The following parameters must be specified for the incinerator:

- Desired control efficiency—This efficiency should be based on requirements dictated by relevant state and federal regulations.
- Combustion chamber outlet temperature—This temperature may also be based on requirements of a regulation or on recommendations developed during regulatory development.
- Desired percent energy recovery—The desired percent energy recovery should be the result of a process optimization in which costs of incinerators with several different levels of energy recovery are estimated and the minimum cost design selected. The

**Table 3.4:** Specification of Sample Problem

Variable	Value
Preheater Inlet Waste Gas Vol Flow Rate, $Q_{w_i}$ scfm	20,000
Preheater Inlet Waste Gas Temp., $T_{w_i}$ , °F	100
Composition	
Benzene Content, ppmv	1000
Methyl Chloride Content, ppmv	1000
Air Content	Balance
Particulate Content	Negligible
Moisture Content	Negligible
Desired Control Efficiency, %	98
Desired Percent Energy Recovery, HR%	70

tradeoff is between the capital cost of the energy recovery equipment and the operating (fuel) cost.

Specifications for the sample problem are given in Table 3.4.

**Step 2 - Verify that the oxygen content of the waste gas exceeds 20%** There must be sufficient oxygen in the waste gas to support the combustion of the waste organics (including VOCs) and the auxiliary fuel, if auxiliary fuel is needed. It may be necessary to add auxiliary air if the oxygen content is less than about 20%. This example is based on streams that contain > 20% oxygen, as shown below:

$$\begin{aligned} \text{Content, Vol. \%} &= 100.0 - \frac{1000}{10^6} \times 100 - \frac{1000}{10^6} \times 1 \\ &= 99.8\% \end{aligned} \quad 3.7$$

$$\begin{aligned} \text{Oxygen Content, \%} &= \text{Air Content} \times 0.209 \\ &= 20.86\% \end{aligned} \quad 3.8$$

**Step 3 - Calculate the LEL and the Percent of the LEL of the gas mixture** Note: If the waste stream contains a significant amount of inerts in addition to the nitrogen associated with the oxygen in air, the calculation of LEL (and UEL) loses meaning since the LEL (and UEL) is measured in mixtures of organic with air only. A complete chemical analysis is necessary to complete the design procedure in such a case.

The example chosen here is typical, in that there is more than one VOC component in the gas stream. An approximate method to calculate the LEL of a mixture of compounds,  $LEL_{mix}$  is given by Grelecki [13] as

$$LEL_{mix} = \left[ \frac{\sum_{j=1}^n x_j}{\left( \sum_{i=1}^n x_i \right) \times LEL_j} \right]^{-1} \quad 3.9$$

where

$$\begin{aligned} x_i &= \text{volume fraction of combustible component } i \\ LEL_j &= \text{lower explosive limits of combustible component } j \text{ (ppmv)} \\ n &= \text{number of combustible components in mixture} \end{aligned}$$

For the example case,

$$\begin{aligned} \sum_{i=1}^n x_i &= (1,000 + 1,000) \times 10^{-6} \\ &= 2,000 \times 10^{-6} \end{aligned} \quad 3.10$$

From standard references [14] or from Appendix 3A,

$$\begin{aligned} LEL_{Bz} &= 14,000 \text{ ppmv for benzene} \\ LEL_{MC} &= 82,500 \text{ ppmv for methyl chloride} \end{aligned}$$

$$LEL_{mix} = \left[ \frac{1000}{2,000 \times 14,000} + \frac{1000}{2,000 \times 82,500} \right]^{-1} = 23,938 \text{ ppmv} \quad 3.11$$

$$LEL_{mix} = \frac{\text{total combustible conc. in mixture}}{LEL_{mix}} \times 100 \quad 3.12$$

$$\frac{2,000}{23,938} \times 100 = 8.4\% \quad 3.13$$

The percent LEL of the mixture is therefore 8.4%. Because this is well below 25%, no dilution air is needed in this example. If the mixture had been above 25% LEL, sufficient dilution air would have been needed to bring the concentration of the mixture to less than 25% to satisfy fire insurance regulations.

**Step 4 - Calculate the volumetric heat of combustion of the waste gas streams,  $(-\Delta h_{c_w})$ , Btu/scf** The energy content of the gas stream, expressed in terms of the heat of combustion, is calculated as follows:

$$(-\Delta h_{c_w}) = \sum_{i=1}^n (-\Delta h_{c_i}) x_i \quad 3.14$$

where

$(-\Delta h_{c_w})$  = heat of combustion of the waste stream (Btu/scf)

$(-\Delta h_{c_i})$  = volumetric heat of combustion of component  $i$  at 25°C (Btu/scf)

$x_i$  = volume fraction of component  $i$  in the waste gas

$n$  = number of combustible components in the waste gas

The heat of combustion that should be used in these calculations is the "lower" heat of combustion, *i.e.*, with gaseous water, rather than liquid water, as a reaction product since water leaves the incinerator in the vapor state. From Appendix 3A or standard references [14,15] with appropriate conversion of units, the volumetric heat of combustion at 25°C for the two components is calculated to be as follows:

$(-\Delta h_{c_{Bz}})$  = 3,475 Btu/scf for benzene

$(-\Delta h_{c_{MC}})$  = 705 Btu/scf for methyl chloride

The compositions specified earlier as ppmv are converted to volume fractions as follows:

$x_{Bz}$  = 1,000 ppmv  $\times 10^{-6}$  =  $10^{-3}$  for benzene

$x_{MC}$  = 1,000 ppmv  $\times 10^{-6}$  =  $10^{-3}$  for methyl chloride

Using these values of heat of combustion and composition, the heat of combustion of the waste gas stream per *standard cubic foot* of incoming gas is



$$\begin{aligned} (-\Delta h_{c_w}) &= (3,475) (10^{-3}) - (705) (10^{-3}) \\ &= 4.18 \text{ Btu/scf} \end{aligned} \quad 3.15$$

Assuming the waste gas is principally air, with a molecular weight of 28.97 and a corresponding density of 0.0739 lb/scf, the heat of combustion per *pound* of incoming waste gas is

$$(-\Delta h_{c_w}) = 56.6 \text{ Btu/lb}$$

The negative heat of combustion, by convention, denotes an exothermic reaction. Also by convention, if one refers to heat of reaction rather than heat of combustion, then a positive value denotes an exothermic reaction.

Empirically, it has been found that 50 Btu/scf roughly corresponds to the LEL of organic/air mixtures. Insurance codes require a value below 25% LEL, which corresponds to about 13 Btu/scf. However, if LEL sensors and monitors are installed, one can incinerate a waste gas with a combustible organic content between 25 and 50% LEL, which corresponds to 13 to 25 Btu/scf.

For catalytic applications the heat of combustion must normally be less than 10 Btu/scf (for VOCs in air) to avoid excessively high temperatures in the catalyst bed. This is, of course, only an approximate guideline and may vary from system to system.

After Step 4, determination of the  $(-\Delta h_{c_w})$  design procedure for thermal and catalytic incinerators is discussed separately, beginning with Step 5 for each type of incinerator.

### 3.4.2 Steps Specific to Thermal Units

Figure 3.1 shows a generic thermal incinerator with the appropriate streams labeled.

**Step 5t - Establish the temperature at which the incinerator will operate** As mentioned in Section 3.2.1, both the reactor temperature and residence time of the waste gas in the reactor determine the level of VOC destruction. In general, state and local regulations specify the required level of destruction that the customer must meet. In this example a destruction efficiency of 98 percent is specified. Studies by Mascone [2,3,4] show that this destruction efficiency can be met in a thermal incinerator operated at a temperature,  $T_R$ , of 1,600°F and a residence time of 0.75 second. (Note: This higher efficiency level is the *minimum* achievable by any new properly designed and operated incinerator. Many incinerators can achieve destruction efficiencies of 99% or higher.)

**Step 6t - Calculate the waste gas temperature at the exit of the preheater** The extent of the heat exchange to be carried out in the preheater is the result of a technical and economic optimization procedure that is not illustrated in this example. As the VOC stream temperature leaving the heat exchanger,  $T_{w_o}$ , increases, the auxiliary fuel requirement decreases, but at the expense of a larger heat exchanger. However, there are several important limits on  $T_{w_o}$ . First,  $T_{w_o}$  must not be close to the ignition temperature of the organic-containing gas to prevent damaging temperature excursions inside the heat exchanger should the gas ignite. Second, for gases containing halogens, sulfur, and phosphorous (or other acid-forming atoms), the flue gas temperature after the heat exchanger,  $T_{f_o}$ , must not drop below the acid dew point. Both limitations limit the amount of heat exchange and thus the maximum value of  $T_{w_o}$ . The calculation of the acid dew point is not simple. It is recommended that vendor guidance be sought to ensure that the dew point is not reached. Condensation of acid gases will result in corrosion of many of the metals used in heat exchangers. As an example, fuel sulfur contents of 1 to 2 percent can give acid dew points of about 200 to 270°F. Increasing the sulfur content to 4 percent can raise the dew to about 290°F. Chlorine and phosphorous have a much smaller effect on acid dew elevation.

With the following assumptions, one can estimate  $T_{w_o}$  using equation 3.2, the definition of fractional energy recovery for a heat exchanger.

- The fractional energy recovery is specified.
- The amount of auxiliary fuel,  $Q_{af}$ , and auxiliary combustion air,  $Q_a$ , are small relative to the waste gas,  $Q_w$ , so that the mass flow rates of gases,  $\rho_w Q_w$  and  $\rho_f Q_f$ , on both sides of the preheater are approximately the same, or

$$\rho_w Q_w \approx \rho_f Q_f$$

- The heat capacities of the gases on both sides of the preheater are approximately the same, regardless of *composition*. This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.
- The *mean* heat capacities above the reference temperature of the gases on both sides of the preheater are approximately the same regardless of *temperature*.

With these assumptions, the equation for fractional energy recovery for a heat exchanger becomes

$$\text{Fractional Energy Recovery} = \frac{T_{w_o} - T_{w_i}}{T_{f_i} - T_{w_i}} \quad 3.16$$

For this example with a fractional energy recovery of 0.70, an incinerator operating temperature,  $T_{fi}$ , of  $1600^{\circ}\text{F}$ , and a waste gas inlet temperature,  $T_{wi}$ , of  $100^{\circ}\text{F}$ , the waste gas temperature at the end of the preheater becomes

$$T_{wo} = 150^{\circ}\text{F}$$

The temperature of the exhaust gas,  $T_{fo}$ , can be determined by an energy balance on the preheater, which, with the same assumptions as used in deriving Equation 3.16 regarding the mass flow rates and average heat capacities of the gases involved, results in the following equation:

$$T_{fi} - T_{fo} = T_{wo} - T_{wi}$$

i.e., the temperature rise in the waste gas is approximately equal to the temperature decrease in the flue gas with which it is exchanged. For this example, this results in

$$T_{fo} = 550^{\circ}\text{F}$$

This value of  $T_{fo}$  should be well above the acid dew point of the flue gas stream.

It should be remembered that  $T_{wo}$  should be well below the ignition temperature of the VOC stream to prevent unwanted temperature excursions in the preheater. This must be verified even if the stream is well below the LEL because flammability limits can be expanded by raising the reactant stream temperature. A sufficiently high preheat temperature,  $T_{wo}$ , could initiate reaction (with heat release) in the preheater. This would ordinarily be detrimental to the materials of construction in the heat exchanger. The one exception is the thermal incinerator of the regenerable type described in Section 3.2. The 95-percent energy recovery, obtainable in regenerable systems would result in this example in a  $T_{wo}$  of  $1,525^{\circ}\text{F}$ . The significant reaction rate that would occur at this temperature in the ceramic packing of the heat exchanger/reactor is by design.

**Step 7t - Calculate the auxiliary fuel requirement,  $Q_{af}$**  Auxiliary fuel will almost invariably be needed for startup of the unit. However, at steady state, if the energy released by combustion of the organics present in the waste stream is sufficient to maintain the reactor temperature ( $1,600^{\circ}\text{F}$  in the example), only a small amount of auxiliary fuel ( $< 5\%$  of the total energy input) is needed to stabilize the flame. In most cases, however, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature.

With the following assumptions, one can estimate  $Q_{af}$  using a mass and energy balance around the combustion chamber and following the principles discussed in Section 3.3, with reference to Figure 3.1.

- The reference temperature,  $T_{ref}$  is taken as the inlet temperature of the auxiliary fuel,  $T_{af}$ .
- No auxiliary air,  $Q_a$ , is required.
- Energy losses,  $H_L$ , are assumed to be 10% of the total energy input to the incinerator above ambient conditions. [16,17] Thus, if the reference temperature is near ambient conditions,

$$H_L = 0.1 \rho_{fi} Q_{fi} C_{pm_{fi}} (T_{fi} - T_{ref}) \quad 3.17$$

- The heat capacities of the waste gases entering and leaving the combustion chamber are approximately the same, regardless of *composition*. This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.
- The *mean* heat capacities above the reference temperature of the waste gases entering and leaving the combustion chamber are approximately the same regardless of *temperature*. Thus the mean heat capacity for the waste gas stream entering or leaving the combustion chamber should be evaluated at the average of  $T_{w_o}$  and  $T_{fi}$ . For air this assumption introduces an error of, at most, 5% over the temperatures of interest.

With these assumptions, the mass and energy balance around the combustion chamber reduces to the following equation:

$$\rho_{w_i} Q_{w_i} [C_{pm_{air}} (1.1 T_{fi} - T_{w_o} - 0.1 T_{ref}) - (-\Delta h_{c_{wo}})] = \rho_{af} Q_{af} [C_{pm_{air}} (1.1 T_{fi} - T_{ref}) - (-\Delta h_{c_{af}})] \quad 3.18$$

Input data for this equation are summarized below:

- The waste stream is essentially air so that

$$\rho_{w_o} = \rho_{w_i} = 0.0739 \text{ lb/scf, air at } 77^\circ\text{F, } 1 \text{ atm.}$$

**Table 3.5:** Summary of Example Problem Variable Valuation

$$T_{ref} = 77^\circ\text{F}$$

Stream	Subscript, $j$	$\rho_j$ lb/scf	$Q_j$ scfm	$C_{pmj}$ Btu/lb °F	$T_j$ °F
IN - Sensible heat					
Auxiliary Air	a	na*	na*	na*	na*
Auxiliary Fuel	af	0.0408	167	**	77
Waste Gas	$w_o$	0.0739	20,000	0.255	1,150
OUT - Sensible Heat					
Waste Stream	$f_i$	0.0739	20,167	0.255	1,600

$(-\Delta h_c)$ , waste gas = 56.6 Btu/lb

$(-\Delta h_c)$ , auxiliary fuel = 21,502 Btu/lb

\*Not applicable.

\*\*Not used because reference temperature is taken equal to auxiliary fuel temperature.

$C_{pm_{air}} = 0.255 \text{ Btu/lb } ^\circ\text{F}$ , the mean heat capacity of air between  $77^\circ\text{F}$  and  $1,375^\circ\text{F}$  (the average temperature of the waste gas entering and leaving the combustion chamber)

- Other input data to Equation 3.18 include:

$$\begin{aligned}
 Q_{w_o} &= Q_{w_i} = 20,000 \text{ scfm} \\
 (-\Delta h_c)_{af} &= 21,502 \text{ Btu/lb, for methane} \\
 T_{af} &= T_{ref} = 77^\circ\text{F, assume ambient conditions} \\
 \rho_{af} &= 0.0408 \text{ lb/ft}^3, \text{ methane at } 77^\circ\text{F, 1 atm.} \\
 T_{fi} &= 1,600^\circ\text{F, Step 5t} \\
 T_{w_o} &= 1,150^\circ\text{F, Step 6t} \\
 (-\Delta h_c)_{w_o} &= 56.6 \text{ Btu/lb. Step 4}
 \end{aligned}$$

substituting the above values into Equation 3.18 results in:

$$Q_{af} = 167 \text{ scfm}$$

**Table 3.6:** Terms in Energy Balance Around Combustor—Example Problem

Stream	Subscript, j	Value, Btu/min
IN - Sensible Heat, $\rho_j Q_j C_{pmj} (T_i - T_{ref})$		
Auxiliary Air	a	0
Waste Gas	$w_o$	404,403
OUT - Sensible Heat, $\rho_j Q_j C_{pmj} (T_i - T_{ref})$		
Waste Stream	$f_i$	578,796
OUT - Losses		
10% of total energy input		57,880
GENERATION -		
Heat of Combustion, $\rho_j Q_j (-\Delta h_{ej})$		
Waste Gas	$w_o$	83,655
Auxiliary Fuel	af	146,506

The values of the parameters in the energy balance are summarized in Table 3.5.

It is instructive to examine the magnitude of the various terms in the energy balance around the combustor for the sample problem. This is done in Table 3.6. The energy balance shown does not quite add to zero due to round-off-error and simplifying assumptions. Table 3.6 shows that the largest inlet term is the sensible heat of the incoming waste gas. The heat of combustion of the organics contained in the waste gas stream is somewhat smaller than that of the auxiliary methane because of the relatively small amount of organics in the waste gas stream. The largest term in the outlet stream is the sensible heat of the outgoing waste stream. The overall energy losses are based on an assumption, but are relatively small. Because the sensible heat contents of the entering and leaving waste stream are so large, it is apparent that energy recovery is an important factor in achieving energy efficiency. In fact, with zero energy recovery in the sample problem, the auxiliary fuel requirements would be 605 scfm, about four times the energy requirements based on 70% energy recovery.

**Step 8t - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame** Only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7t is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used, and the amount of energy recovery, specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5t).

This check is made by calculating 5% of the total energy input to the incinerator and comparing it with the auxiliary fuel energy input. The total energy input is given as follows:

$$\text{Total Energy Input} = Q_{fi} C_{pm_{fi}} (T_{fi} - T_{ref}) \quad 3.19$$

$$\text{Auxiliary Fuel Energy Input} = Q_{af} \left( \frac{\Delta H}{c_{af}} \right) \quad 3.20$$

The auxiliary fuel used in the design,  $Q_{af}$  should be the larger of 5% of the total energy input (28,900 Btu/min.) and the auxiliary fuel energy input (146,500 Btu/min.). The auxiliary fuel used easily meets this criterion.

**Step 9t - Calculate the total volumetric flow rate of gas through the incinerator,  $Q_{fi}$**  The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate,  $Q_{fi}$  and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the incinerator.

$$\begin{aligned} Q_{fi} &= Q_{w_o} + Q_a + Q_{af} \\ &= 20,000 + 0 + 167 \\ &= 20,167 \text{ scfm} \end{aligned}$$

This result conforms with the assumptions stated in Step 6t, *i.e.*, the mass (and volume) flow rates on both sides of the preheater are approximately equal. Finally, it must be emphasized that steps 5t to 9t apply to thermal *recuperative* incinerators, only. To calculate the auxiliary fuel requirements for other types of thermal incinerators (*e.g.*, regenerative), a different procedure must be used. (See Appendix 3B.)

### 3.4.3 Steps Specific to Catalytic Units

Figure 3.3 shows a generic catalytic incinerator with the appropriate streams labeled. The approach used in the calculations on the catalytic incinerator is somewhat different than that used in the thermal incinerator. This difference arises because of additional constraints which are placed on the catalytic incinerator. These constraints are as follows:

- The desired catalyst bed outlet temperature is typically 700 to 900°F. The maximum temperature to which the catalyst bed can be exposed continuously is limited to about 1,200°F. Therefore, the combustible content of the waste gas is limited, and the amount of heat exchange that occurs in the primary heat exchanger may be limited.
- The inlet temperature to the catalyst bed itself must be above the catalytic ignition temperature required to give the desired destruction efficiency in the incinerator. Therefore, the combustible content of the waste gas is further limited to that which,

when combusted, will raise the temperature in the catalyst bed no more than the  $\Delta T$  between the required reactor bed inlet temperature, and the desired reactor bed outlet temperature.

- Auxiliary fuel, in combination with the preheat from the primary heat exchanger, is used to preheat the waste gas to the reactor inlet temperature. A minimum amount of auxiliary fuel ( $< 5\%$  of the total energy input) must be used to stabilize the flame in the preheat combustion chamber. This has the effect of further limiting the combustible content of the waste gas stream and the amount of heat exchange permissible in the primary heat exchanger.

The steps outlined below represent one approach to recognizing these constraints and incorporating them into the calculation procedures.

**Step 5c - Establish the desired outlet temperature of the catalyst bed,  $T_{fi}$**  The energy released by the oxidation of the VOCs in the catalyst bed will raise the temperature of the gases by an amount,  $\Delta T$ , as the gases pass through the catalyst bed. An outlet temperature from the catalyst, and thus from the reactor, must be specified that will ensure the desired level of destruction of the VOC stream. As in thermal incinerators, this value varies from compound to compound and also varies from catalyst to catalyst. Final design of the incinerator should be done by firms with experience in incinerator design. Guidelines given by Combustion Engineering [12] indicate that values from 300 to 900°F result in destruction efficiencies between 90 and 95 percent. To prevent deactivation of the catalyst a maximum bed temperature of 1,200°F should not be exceeded. In the example problem the catalyst outlet temperature,  $T_{fi}$  is selected to be 900°F.

**Step 6c - Calculate the waste gas temperature at the exit of the preheater (primary) heat exchanger** The waste gas temperature at the exit of the primary heat exchanger is estimated in the same manner as for the thermal incinerator. The equation for fractional energy recovery Equation 3.16, is used, with the same assumptions as used for the thermal incinerator. For the example problem with a fractional energy recovery of 0.70, a catalyst bed outlet temperature,  $T_{fi}$ , of 900°F, and a waste gas inlet temperature,  $T_{wi}$ , of 100°F, the gas temperature at the exit of the preheater becomes

$$T_{wo} = 660^\circ\text{F}$$

The same considerations regarding the closeness of the temperature of the exhaust gas,  $T_{fe}$  to its dew point apply to the catalytic incinerator as they did to the thermal incinerator.

**Step 7c - Calculate the auxiliary fuel requirement,  $Q_{af}$**  The auxiliary fuel requirement,  $Q_{af}$  is calculated by making mass and energy balances around the preheater combustion chamber and the catalyst chamber. The auxiliary fuel requirement calculated in this manner must be checked to insure that it falls within the constraints imposed by design considerations of the catalytic incinerator. These constraints are as follows:



- The auxiliary fuel requirement must be positive. A negative fuel requirement indicates that the heat of combustion of the waste gas,  $(-\Delta h_c)$ , is too high for the fractional energy recovery in the primary heat exchanger that was selected.
- The auxiliary fuel amount must be high enough to provide a stable flame in the preheater combustion chamber (See Step 8c below).

An energy balance around the preheater combustion chamber and the catalyst chamber, taken together, results in Equation 3.18, the same equation used in the thermal incinerator calculations. The input data for Equation 3.18 for the catalytic incinerator example problem are summarized below:

- The waste stream is essentially air so that

$$\begin{aligned}\rho_{w_o} &= \rho_{w_i} = 0.0739 \text{ lb/scf, air at } 77^\circ\text{F, 1 atm} \\ C_{pm_{air}} &= 0.248 \text{ Btu/lb } ^\circ\text{F, the mean heat capacity of air between } 77^\circ\text{F and } 780^\circ\text{F (the average of the preheater exit and catalyst bed outlet temperatures)}\end{aligned}$$

- Other input data to Equation 3.18 include

$$\begin{aligned}Q_{w_o} &= Q_{w_i} = 20,000 \text{ scfm} \\ (\Delta h_{c_{af}}) &= 21,502 \text{ Btu/lb, for methane} \\ T_{af} &= 77^\circ\text{F, assume ambient conditions} \\ \rho_{af} &= 0.0408 \text{ lb/ft}^3, \text{ methane at } 77^\circ\text{F, 1 atm.} \\ T_{fi} &= 900^\circ\text{F, from Step 5c} \\ T_{w_o} &= 660^\circ\text{F, from Step 6c} \\ (\Delta h_{c_w}) &= 56.6 \text{ Btu/lb, from Step 4}\end{aligned}$$

Substituting the above values into Equation 3.18 results in

$$Q_{af} = 40 \text{ scfm}$$

If the outlet temperature of the catalyst bed,  $T_{fi}$ , is  $800^\circ\text{F}$ , then  $Q_{af}$  decreases to  $-6.7 \text{ scfm}$ . In other words, no auxiliary fuel would, theoretically, be required at this bed temperature. However, as discussed above in Step 8t, a certain quantity of auxiliary fuel would be required to maintain burner stability.

At 70% energy recovery and  $900^\circ\text{F}$  outlet catalyst bed temperature, a waste gas with a heat of combustion,  $(-\Delta h_{c_w})$ , of about  $79.9 \text{ Btu/lb}$  would cause the auxiliary fuel requirement,  $Q_{af}$ , to become negative, indicating the catalyst bed would exceed  $900^\circ\text{F}$ . At 70% energy recovery and  $800^\circ\text{F}$  outlet catalyst bed temperature, this same result occurs with a  $(-\Delta h_{c_w})$  of  $52.7 \text{ Btu/lb}$ .

Both of these heats of combustion are relatively low for typical waste gases. These results are, of course, dependent on the assumption of energy losses from the combustion chamber. The lower the energy losses, the lower the allowable waste gas heat of combustion before overheating occurs in the catalyst bed.

**Step 8c - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame** Only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7c is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used and the amount of energy recovery specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5c).

This check is made in the same manner as that in Step 8t of the thermal incinerator calculation. The results of this check indicate that the auxiliary fuel requirement is more than sufficient to stabilize the burner flame.

**Step 9c - Estimate the inlet temperature to the catalyst bed,  $T_{r_i}$**  The inlet temperature to the catalyst bed must be calculated to ensure that the inlet temperature is above that necessary to ignite the combustible organic compounds in the catalyst that was selected for use.

The inlet temperature to the catalyst bed,  $T_{r_i}$ , should be such that, when the temperature rise through the catalyst bed,  $\Delta T$ , is added to it, the resulting temperature is  $T_{f_i}$ , 900°F. Thus,

$$\Delta T = T_{f_i} - T_{r_i} \quad (3.21)$$

The value of  $\Delta T$  is determined by an energy balance around the preheater portion of the combustor. The preheater is required to heat the gases up to the catalyst bed inlet temperature using auxiliary fuel.<sup>1</sup> This energy balance is made with the assumptions made earlier in deriving Equation 3.18 and further assuming that only auxiliary fuel is combusted in the preheater portion. The resulting equation is very similar to Equation 3.18 except that (1) the terms with an  $f_i$  subscript become terms with  $r_i$  subscripts to denote a catalytic reactor inlet stream rather than a combustor outlet (flue gas inlet to the primary heat exchanger) and (2) the term for combustion of the waste gas organics does not appear. The resulting equation is as follows:

---

<sup>1</sup> At equilibrium, the temperature of the catalyst bed is maintained without requiring auxiliary fuel.

$$\rho_{af} Q_{af} = \frac{\rho_{w_o} Q_{w_o} [C_{pm_{air}} (1.1 T_{r_i} - T_{w_o} - 0.1 T_{ref})]}{(\Delta h_{c_{af}}) - 1.1 C_{pm_{air}} (T_{r_i} - T_{ref})} \quad 3.22$$

This equation may be rearranged to solve for  $T_{r_i}$  explicitly. This produces an equation that is somewhat complex and non-intuitive.

$$T_{r_i} = \frac{\rho_{af} Q_{af} [(\Delta h_{c_{af}}) - 1.1 C_{pm_{air}} T_{ref}] - \rho_{w_o} Q_{w_o} C_{pm_{air}} (T_{w_o} - 0.1 T_{ref})}{1.1 C_{pm_{air}} (\rho_{af} Q_{af} - \rho_{w_o} Q_{w_o})} \quad 3.23$$

After substituting the example problem parameters into Equation 3.23, we obtain a value for  $T_{r_i}$  of 693°F. Based on ignition temperatures shown in Table 3.2, this reactor inlet temperature should be satisfactory. Prior to a more definitive design, the ignition temperatures for the specific chemicals should be verified.

The temperature rise across the catalyst bed is thus (900 - 693) or 207°F. These temperatures are somewhat sensitive to the assumption for energy losses from the combustor. The assumption for energy losses is perhaps somewhat conservative, *i.e.*, it causes a larger  $Q_{af}$  to be estimated than would a less conservative assumption, and becomes more conservative as the combustor size and insulation are increased.

**Step 10c - Calculate the total volumetric flow rate of gas through the incinerator,  $Q_{fi}$**  The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate,  $Q_{fi}$ , and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the incinerator.

$$\begin{aligned} Q_{fi} &= Q_{w_o} + Q_a + Q_{af} \\ &= 20,000 + 0 + 40 \\ &= 20,040 \text{ scfm} \end{aligned}$$

**Step 11c - Calculate the volume of catalyst in the catalyst bed** If the volumetric flow rate of gas through the catalyst bed,  $Q_{fi}$ , and the nominal residence time (reciprocal space velocity) in the catalyst bed are known, then the volume of catalyst can be estimated. There exists complex set of relationships between the catalyst volume and geometry, overall pressure drop across the catalyst, conversion of the oxidizable components in the gas, gas temperature, and the reaction rate. These relationships are dependent on the catalyst and the type of compound being oxidized. It is beyond the scope of this *Manual* to discuss these relationships, even in an

approximate way. For the purposes of cost estimation, the space velocity,  $\phi$  in reciprocal time units, necessary to achieve the required level of destruction can be used to approximate the catalyst volume

requirement. The space velocity is defined as

$$\phi = \frac{Q_{fi}}{V_{cat}}$$

where

$V_{cat}$  = Overall bulk volume of the catalyst bed, including interparticle voids (ft<sup>3</sup>)

By petro-chemical industry convention, the space velocity is computed at the conditions of 60°F (not 77°F) and 1 atm. The volumetric flow rate,  $Q_{fi}$  must be corrected to these conditions. The proper space velocity to achieve a desired level of conversion is based on experimental data for the system involved. For precious metal monolithic catalysts, the space velocity generally lies between 10,000 h<sup>-1</sup> and 60,000 h<sup>-1</sup>. (Base metal catalysts operate at lower space velocities, ranging from 5,000 to 15,000 h<sup>-1</sup>.) [10]

For the example, using a space velocity of 30,000 h<sup>-1</sup> or 500 min<sup>-1</sup>, and using  $Q_{fi}$  at 60°F,

$$\begin{aligned} Q_{fi} \text{ at } 60^\circ\text{F} &= 20,040 \left( \frac{60 - 460}{77 - 460} \right) \\ &= 19,400 \text{ ft}^3/\text{min} \\ V_{cat} &= \frac{19,400 \text{ ft}^3/\text{min}}{500 \text{ min}^{-1}} \\ &= 39 \text{ ft}^3 \end{aligned}$$

There are a number of catalyst bed parameters, such as catalyst configuration and bed design, that are not significant for study type cost estimates. Accordingly, design of these factors is not discussed here.

**Table 3.7:** Scope of Cost Correlations

<b>Incinerator Type</b>	<b>Total (Flue) Gas Flowrate, scfm</b>	<b>Figure Number</b>
Thermal - Recuperative	500 <sup>a</sup> -50,000	3.4
Thermal - Regenerative	10,000-100,000	3.5
Fixed-Bed Catalytic	2,000-50,000	3.6
Fluid-Bed Catalytic	2,000-25,000	3.7

<sup>a</sup> Although Figure 3.4 covers the 1,000 to 50,000 scfm range, the correlation is valid for the 500 to 50,000 scfm range.

## 3.5 Cost Analysis

This section presents procedures and data for estimating capital and annual costs for four types of incinerators: (1) thermal-recuperative, (2) thermal regenerative, (3) fixed-bed catalytic, and (4) fluid-bed catalytic.

### 3.5.1 Estimating Total Capital Investment

Total capital investment, TCI, includes the equipment cost, EC, for the incinerator itself, the cost of auxiliary equipment (e.g., ductwork), all direct and indirect installation costs, and costs for buildings, site preparation, offsite facilities, land, and working capital. However, the last five costs usually do not apply to incinerators. (See Chapter 2 of this *Manual* for a detailed description of the elements comprising the TCI)

#### 3.5.1.1 Equipment Costs, EC

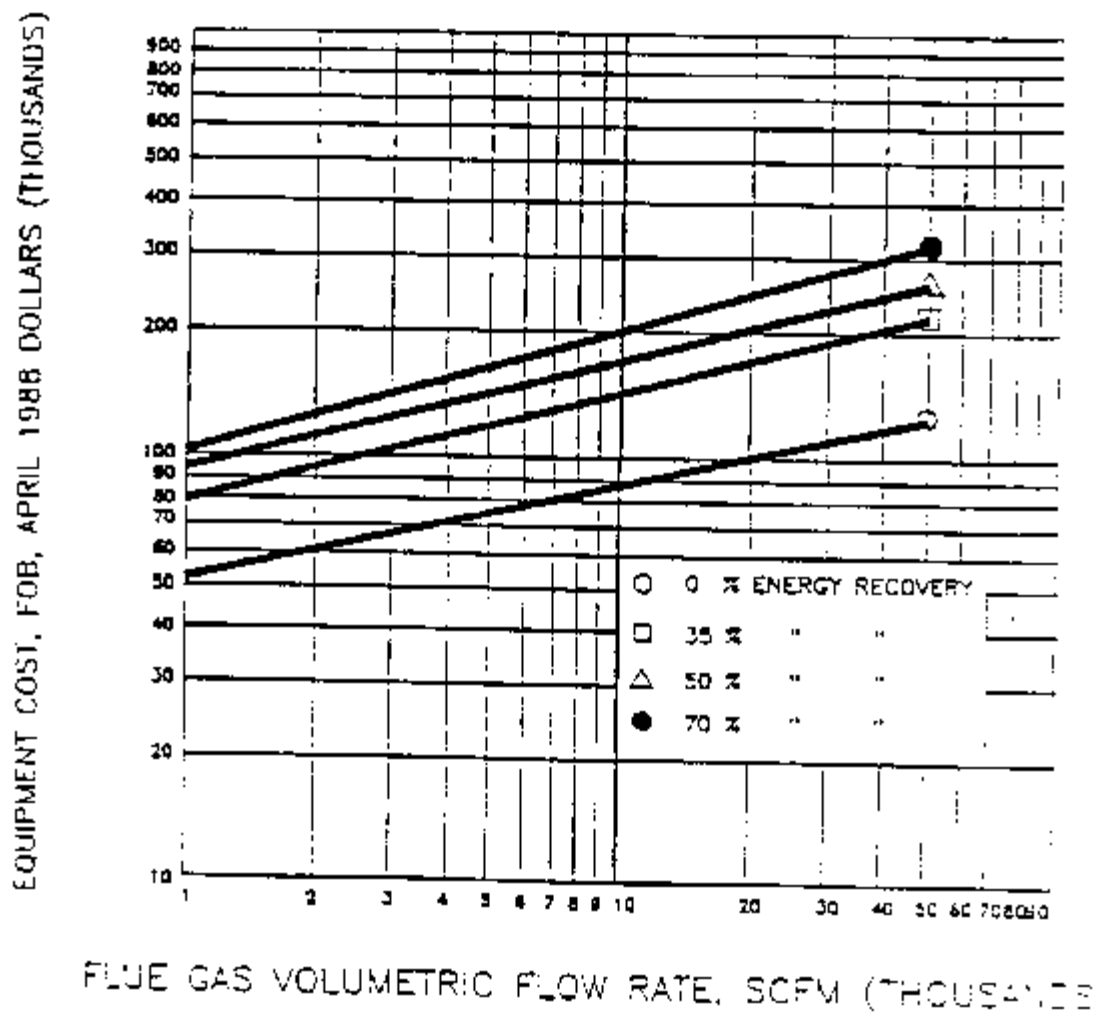
As discussed in Section 3.2.3, the equipment costs, EC, given in this chapter apply to packaged incinerators, except for regenerative incinerators. For regenerative incinerators, the costs apply to field-erected units. The EC typically includes all flange-to-flange equipment needed to oxidize the waste gas, including the auxiliary burners, combustion chamber, catalyst, primary heat exchanger (except for the "zero heat recovery" cases), weathertight housing and insulation, fan, flow and temperature control systems, a short stack, and structural supports. Smaller units, e.g., typically less than 20,000 scfm, are typically preassembled skid-mounted [18]. The various available incineration systems are presented in four groups delineated according to their similarity of design. These groups are outlined in Table 3.7. With the exception of regenerative thermal and fluid-bed catalytic incinerators, the maximum size for which costs are given is 50,000 scfm. Although larger units of each technology can be built, applications are rare at flow rates above 50,000 scfm. Regenerative thermal incinerator costs are provided for flow rates from 10,000 to 100,000 scfm. Fluid-bed catalytic incinerator costs are provided for flow rates from 2,000 to 25,000 scfm.

The cost curves are least-squares regressions of cost data provided by different vendors. It must be kept in mind that even for a given incineration technology, design and manufacturing

procedures vary from vendor to vendor, so that costs may vary. As always, once the study estimate is completed, it is recommended that more than one vendor be solicited for a more detailed cost estimate.

The additional expense of acid gas clean-up or particulate control is not treated in this section. The equipment cost of a gas absorber to remove any acid gases formed in the incinerator can be quite large, sometimes exceeding the equipment cost of the incinerator itself even for simple packed tower scrubbers [19]. For more complex absorbers that include venturi scrubbers instead of, or in addition to, packed beds, the cost of the scrubber alone may be up to 4 times that of the incinerator [11]. These more complex absorbers are sometimes necessary when particulates, in addition to acid gases, must be removed from the flue gas. (Note: Chapter 9 of the *Manual* provides data and procedures for sizing and costing gas absorbers.)

**Thermal Incinerators** Among the thermal units, the direct flame (0% energy recovery) and recuperative systems are treated together because the various levels of energy recovery are achieved simply by adding heat exchanger surface area. Costs for these units were provided by several vendors [12,20,21]. The EC of these units are given as a function of *total* volumetric throughput,  $Q_{tot}$ , in *scfm*. " $Q_{tot}$ ", is the total volume of the gaseous compounds exiting the combustion chamber; it is identical to the term, " $Q_{\dot{g}}$ " used in Figures 3.1 and 3.3. This includes the combustion products, nitrogen, unburned fuel and organics, and other constituents. (See Figure 3.4). Note that costs are given free on board (F.O.B.) in April



**Figure 3.4.** Equipment Costs of Thermal Incinerators, Recuperative

1988 dollars\*. Based on a least-squares regression analysis, a log-log relationship between throughput and EC was found for a given level of energy recovery (HR) over the flow rate range from 500 to 50,000 scfm. These relationships are as follows:

$$EC = 10294 Q_{tot}^{0.2355} \quad HR = 0\% \quad (3.24)$$

$$EC = 3149 Q_{tot}^{0.2609} \quad HR = 35\% \quad (3.25)$$

$$EC = 7056 Q_{tot}^{0.2502} \quad HR = 50\% \quad (3.26)$$

$$EC = 1342 Q_{tot}^{0.2500} \quad HR = 70\% \quad (3.27)$$

The regenerative (or excess enthalpy) systems provide up to 95 percent heat recovery at the expense of higher capital costs. Their unique design [22,23], which combines the heat exchanger and reactor, is substantially different from traditional thermal units and is therefore treated separately in Figure 3.5. The ECs of these systems are given as an approximately linear function of total flow rate over a 10,000 to 100,000 scfm range by the following equation:

$$EC = 2.204 \times 10^5 - 11.57 Q_{tot} \quad 3.28$$

Again, the higher capital costs of these units can be substantially offset by the substantial savings in auxiliary fuel costs.

**Catalytic Incinerator** The EC for a catalytic incinerator is a function of the type of catalyst contacting pattern used and the total gas flow rate,  $Q_{tot}$ , for a given level of energy recovery. There are three types of contacting configurations used in catalytic systems: fixed-bed, catalytic monolith, and fluid-bed. The EC for the first two are generally comparable and are given in Figure 3.6. The data provided by several vendors [12,20,21,24] exhibited curvilinear relationships with  $Q_{tot}$  for each of the energy recovery rates. Least squares regressions of the data yielded the following correlations for total flow rates between 2,000 and 50,000 scfm:

$$EC = 1105 Q_{tot}^{0.5471} \quad HR = 0\% \quad (3.29)$$

$$EC = 3623 Q_{tot}^{0.4189} \quad HR = 35\% \quad (3.30)$$

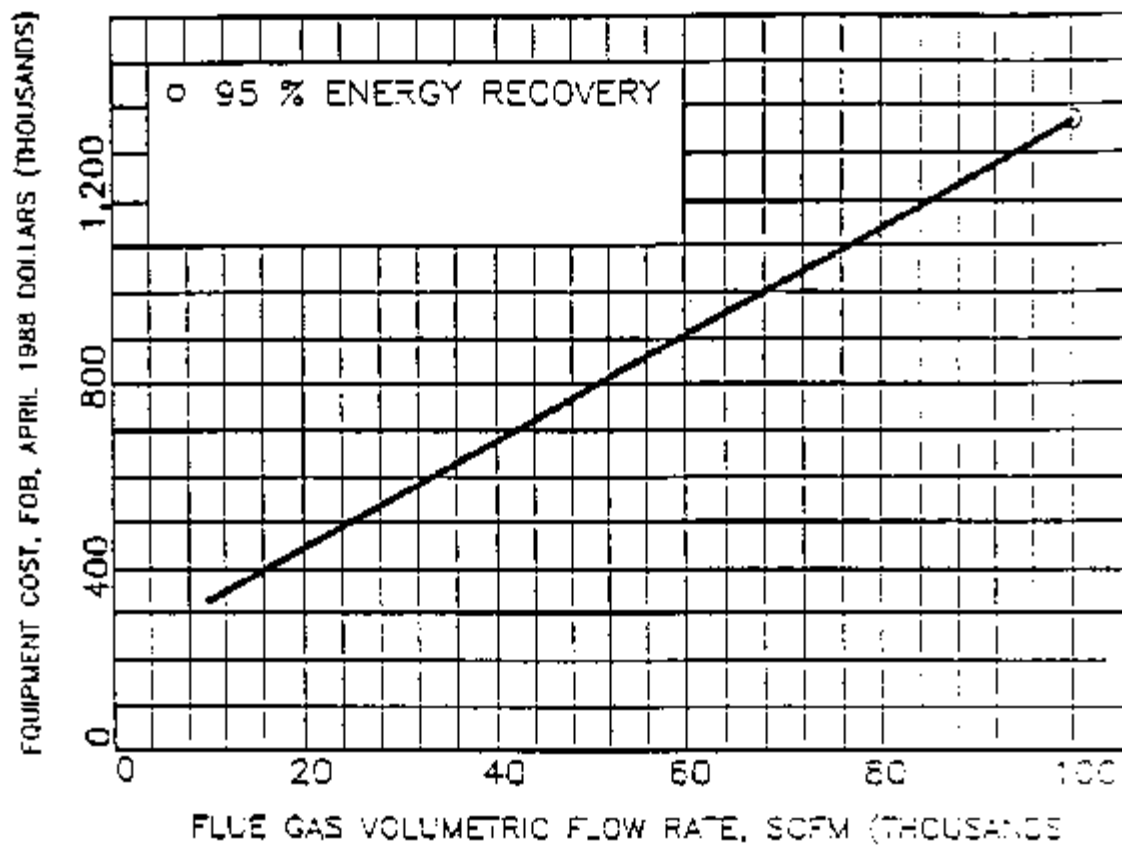
$$EC = 1215 Q_{tot}^{0.5575} \quad HR = 50\% \quad (3.31)$$

$$EC = 1443 Q_{tot}^{0.5527} \quad HR = 70\% \quad (3.32)$$

---

\*For information on escalating these and the other incinerator prices to more current dollars, refer to the EPA report *Escalation Indexes for Air Pollution Control Costs* and updates thereto, all of which are installed on the OAQPS Technology Transfer Network (CTC Bulletin Board).





**Figure 3.5.** Equipment Costs of Thermal Incinerators, Regenerative

Fluid-bed catalytic incinerators afford certain advantages over fixed-bed catalyst units in that they tolerate waste streams with (1) higher heating values, (2) particulate contents, and (3) chlorinated species. For this enhanced flexibility of feed streams, a higher capital cost is incurred, as indicated by the EC shown in Figure 3.7. The data shown were provided by vendors [11,19] and exhibited a linear relationship over the range of flow rates from 2,000 to 25,000 *scfm*. They can be approximated by the following equations:

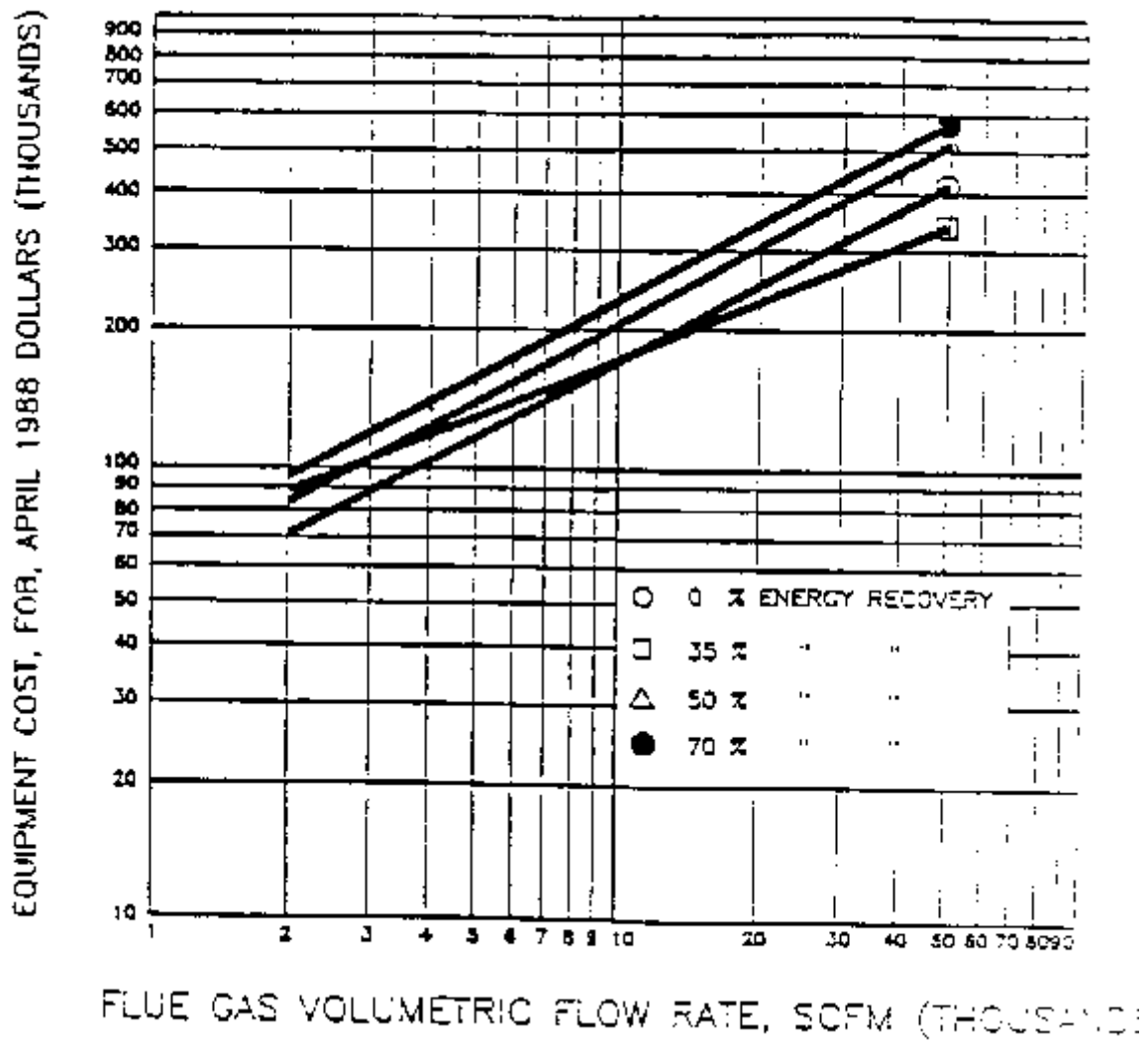
$$EC = 8.48 \times 10^4 - 13.2Q_{tot} \quad HR = 0\% \quad (3.33)$$

$$EC = 8.84 \times 10^4 - 14.6Q_{tot} \quad HR = 35\% \quad (3.34)$$

$$EC = 8.66 \times 10^4 - 15.8Q_{tot} \quad HR = 50\% \quad (3.35)$$

$$EC = 8.39 \times 10^4 - 19.2Q_{tot} \quad HR = 70\% \quad (3.36)$$





**Figure 3.6.** Equipment Costs of Catalytic Incinerators, Fixed-Bed

A comparison of the thermal, catalytic fixed-bed, and catalytic fluid-bed systems with 50 percent energy recovery is shown in Figure 3.8.

### **3.5.1.2 Installation Costs**

As explained in Chapter 2, the purchased equipment cost, PEC, is calculated by taking the sum of the EC and the cost of auxiliary equipment (e.g., ductwork), taxes, freight, and instrumentation. Average values of direct and indirect installation factors [25] to be applied to the PEC are given in Table 3.8 for both recuperative thermal and fixed- and fluid-bed catalytic incinerators.

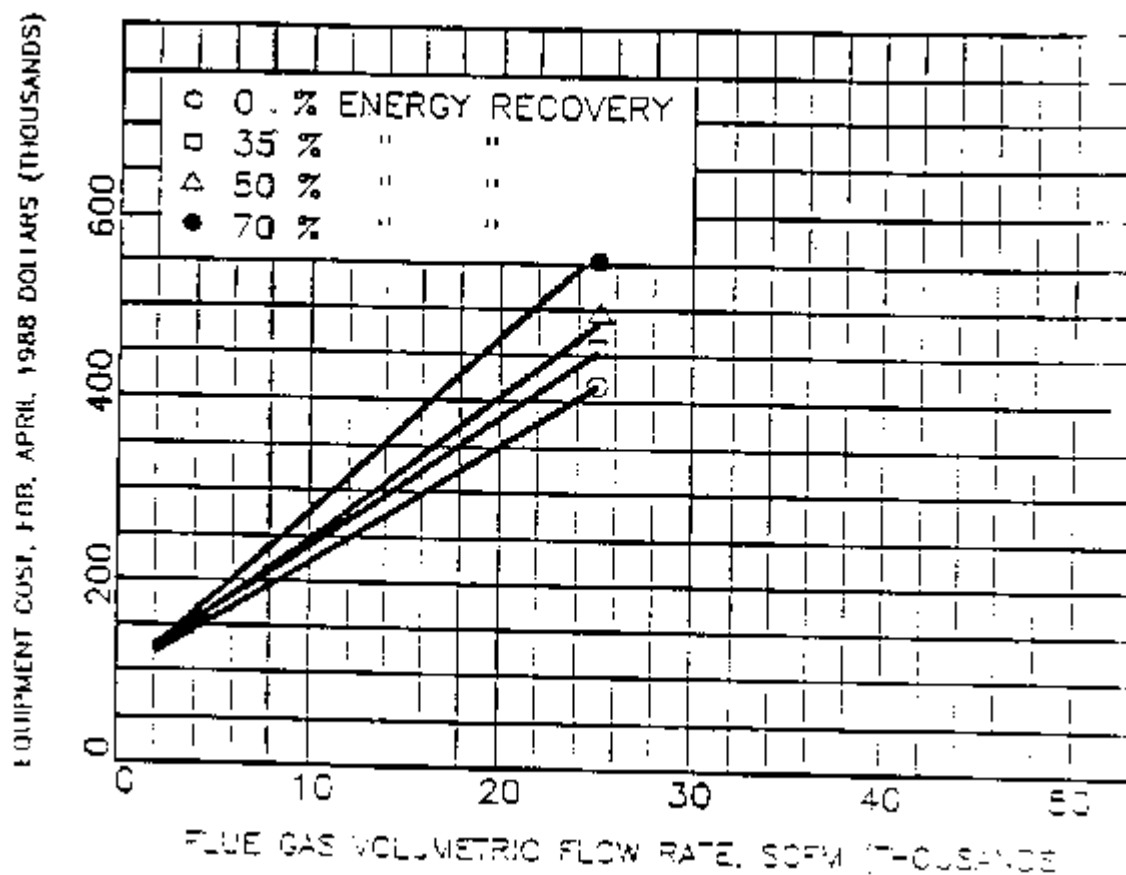
Table 3.9 shows the itemized installation costs that are obtained when these installation factors are applied to the PECs for the example incinerators. Depending on the site conditions, the installation costs for a given incinerator could deviate significantly from costs generated by these average factors. Vataavuk and Neveril [25] provide some guidelines for adjusting the average installation factors to account for other-than-average installation conditions. For units handling total gas flow rates lower than 20,000 scfm the installation costs are minimal, amounting normally to only utility tie-ins (electrical and, if necessary, combustion or dilution air). The installation costs for these smaller incinerators would be 20 to 25 % of the PEC. Smaller units may be installed on the roofs of manufacturing buildings rather than at ground level. In such cases the installation factors could be as high as (or higher than) the factors shown in Table 3.8, even though the units would be "packaged".

## **3.5.2 Estimating Total Annual Cost**

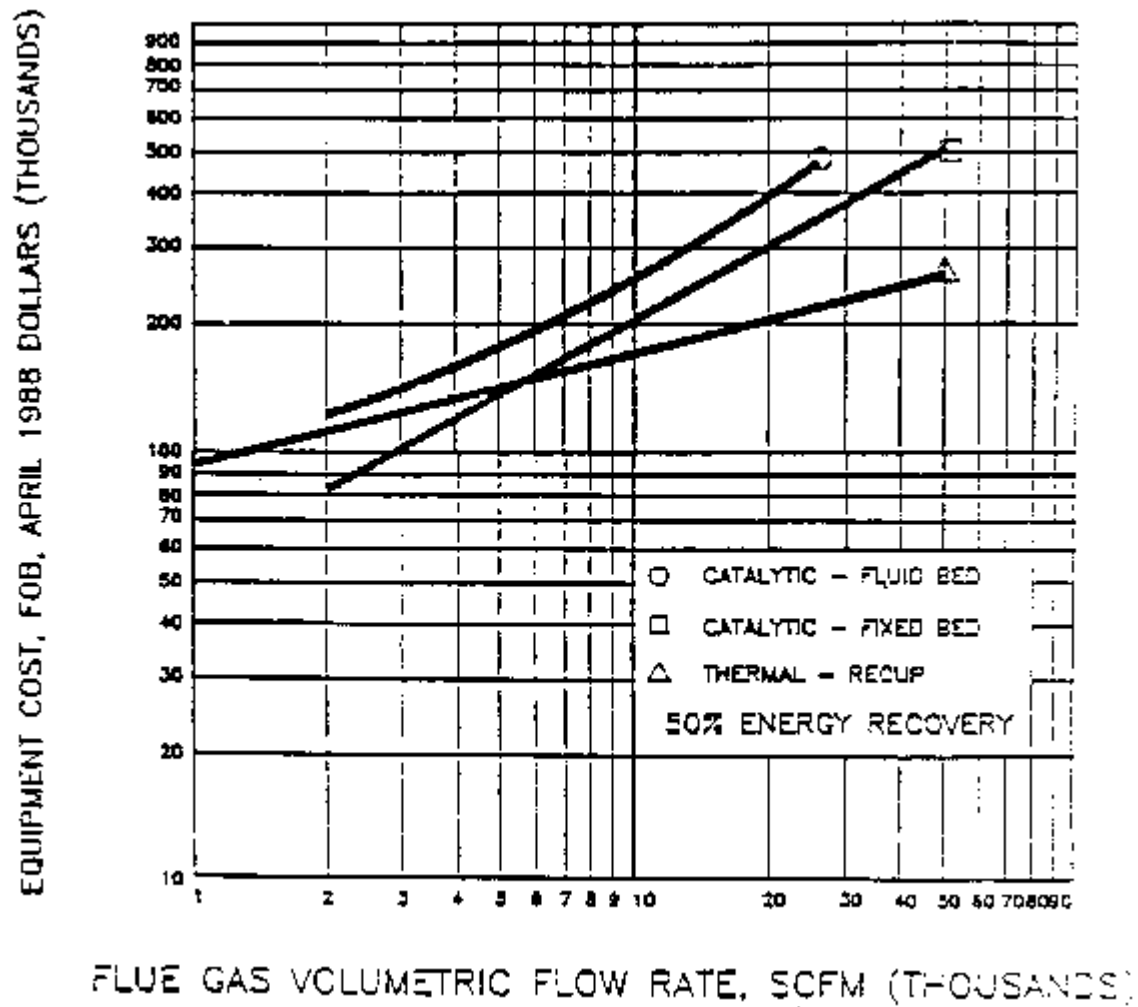
The total annual cost (TAC) is the sum of the direct and indirect annual costs. The TAC for both example systems is given in Table 3.10, along with suggested factors for calculating them.

### **3.5.2.1 Direct Annual Costs**

Direct annual costs for incinerators include labor (operating and supervisory), maintenance (labor and materials), fuel, electricity, and (in catalytic units) replacement catalyst. For thermal and catalytic units, the fuel usage rate is calculated as shown in Sections 3.4.2 and 3.4.3, respectively where natural gas (methane) is assumed to be the fuel. (Other fuels could be used for the thermal units.)



**Figure 3.7.** Equipment Costs of Catalytic Incinerators, Fluid-Bed



**Figure 3.8:** Equipment Costs Comparison of Incinerator Types

**Table 3.8.** Capital Cost Factors for Thermal and Catalytic Incinerators<sup>a</sup>

<b>Cost Item</b>	<b>Factor</b>
<u>Direct Costs</u>	
Purchased equipment costs	
Incinerator (EC) + auxiliary equipment <sup>b</sup>	As estimated, A
Instrumentation <sup>c</sup>	0.10 A
Sales taxes	0.03 A
Freight	<u>0.05 A</u>
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.08 B
Handling & erection	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation for ductwork <sup>d</sup>	0.01 B
Painting	<u>0.01 B</u>
Direct installation cost	0.30 B
Site preparation	As required, SP
Buildings	<u>As required, Bldg</u>
Total Direct Cost, DC	1.30 B + SP + Bldg.
<u>Indirect Costs (installation)</u>	
Engineering	0.10 B
Construction and field expenses	0.05 B
Contractor fees	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingencies	<u>0.03 B</u>
Total Indirect Cost, IC	<u>0.31 B</u>
Total Capital Investment = DC + IC	1.61 B + SP + Bldg.

<sup>a</sup>Reference [25]<sup>b</sup>Ductwork and any other equipment normally not included with unit furnished by incinerator vendor.<sup>c</sup>Instrumentation and controls often furnished with the incinerator, and those often included in the EC.

<sup>d</sup>If ductwork dimensions have been established, cost may be estimated based on \$10 to \$12/ft<sup>2</sup> of surface for fluid application. (Alternatively, refer to Chapter 10 of this Manual.  
 Fan housings and stacks may also be insulated.





**Table 3.9.** Capital Costs for Thermal and Catalytic Incinerators  
Example Problem

Cost Item	Cost, \$	
	Thermal- Recuperative	Fluid-Bed Catalytic
<b>Direct Costs</b>		
Purchased equipment costs		
Incinerator (EC)	\$254,200	\$468,200
Auxiliary equipment*	—	—
Sum = A	\$254,200	\$468,200
Instrumentation, 0.1A	25,400	46,800
Sales taxes, 0.03A	7,630	14,000
Freight, 0.05A	12,700	23,400
Purchased equipment cost, B	\$300,000	\$552,400
Direct installation costs		
Foundation and supports, 0.08B	24,000	44,200
Handling and erection, 0.14B	42,000	77,300
Electrical, 0.04B	12,000	22,100
Piping, 0.02B	6,000	11,000
Insulation (for ductwork), 0.01B	3,000	5,520
Painting, 0.01B	3,000	5,520
Direct installation cost	\$90,000	\$165,500
Site preparation*	—	—
Buildings*	—	—
Total Direct Cost	\$390,000	\$718,000
<b>Indirect Costs (Installation)</b>		
Engineering, 0.10B	30,000	55,200
Construction and field expenses, 0.05B	15,000	27,600
Contractor fees, 0.10B	30,000	55,200
Start-up, 0.02B	6,000	11,000
Performance test, 0.01B	3,000	5,520
Contingencies, 0.03B	9,000	16,600
Total Indirect Cost	\$93,000	\$171,100
Total Capital Investment (rounded)	\$483,000	\$889,000

**Table 3.10.** Annual Costs for Thermal and Catalytic Incinerators  
Example Problem

Cost Item	Suggested Factor	Unit Cost <sup>a</sup>	Thermal	Fluid-Bed Catalyst
<u>Direct Annual Costs<sup>b</sup>, DC</u>				
Operating Labor				
Operator	0.5 h/shift	\$12.95/h	6,480	6,480
Supervisor	15% of operator	—	972	972
Operating Materials	—			
Maintenance				
Labor	0.5 h/shift	\$14.25/h	7,130	7,130
Material	100% of maint. labor	—	7,130	7,130
Catalyst replacement	100% of catalyst replaced ea. 2 yr.	\$650/ft <sup>3</sup> for metal oxide	0	15,100
Utilities				
Natural Gas	—	\$3.30/kft <sup>3</sup>	264,500	63,400
Electricity	—	\$0.059/kWh	<u>36,500</u>	<u>44,200</u>
Total DC			\$321,200	\$144,400
<u>Indirect Annual Costs, IC</u>				
Overhead	60% of sum of operating, supr., & maint. labor & maint. materials.	—	13,000	13,000
Administrative charges	2% TCI	—	9,650	17,800
Property taxes	1% TCI	—	4,830	8,900
Insurance	1% TCI	—	4,830	8,900
Capital recovery <sup>c</sup>	CRF [TCI- 1.08 (Cat. Cost)]	—	68,800	122,700
			<u>\$101,100</u>	<u>\$171,300</u>
Total Annual Cost (rounded)			<u>\$422,000</u>	<u>\$316,000</u>

<sup>a</sup>1988 dollars

<sup>b</sup>Assumes 8,000 h/yr

<sup>c</sup>The capital recovery cost factor, CRF, is a function of the catalyst or equipment life (typically, 2 and 10 years, respectively) and the opportunity cost of the capital (i.e., interest rate). For example, for a 10-year equipment life and a 7% interest rate, CRF = 0.1424.

**Table 3.11.** Typical Pressure Drop Across Selected Equipment

Equipment Type	Energy Recovery, %	$\Delta P$ , in. H <sub>2</sub> O
Thermal Incinerators	0	4
Catalytic Fixed-bed Incinerators	0	6
Catalytic Fluid-bed Incinerators	0	6-10
Heat Exchangers	35	4
" "	50	8
" "	70	15

The electricity costs are primarily associated with the fan needed to move the gas through the incinerator. The power (in kilowatts) needed to move a given inlet volumetric flow rate of air ( $Q_{wi}$  per Sections 3.4.2 and 3.4.3) at a total flange-to-flange pressure drop of  $\Delta P$  inches of water and combined motor/fan efficiency,  $\epsilon$ , is adapted from Equation 2.7, as follows:

$$Power_{fan} = \frac{1.17 \times 10^{-4} Q_{wi} \Delta P}{\epsilon} \quad (3.37)$$

Fan efficiencies vary from 40 to 70 percent [15] while motor efficiencies are usually 90 percent.

The total pressure drop across an incinerator system depends on the number and types of equipment elements included in the system and on design considerations. The estimation of actual pressure drop requirements involves complex calculations based on the specific system's waste gas and flue gas conditions and equipment used. For the purposes of this section, however, the approximate values shown in Table 3.11 can be used.

For the example cases, we will assume 8,000 hours per year operation and a 60% efficiency for the fan and motor together. Using pressure drops of 4 and 8 inches of water, respectively, for the thermal and fluid-bed catalytic incinerators\*, and adding the pressure drop of 15 inches of water for 70% heat recovery, the fan power requirements can be calculated as follows:

Thermal Incinerator\*\*

$$Power_{fan} = \frac{1.17 \times 10^{-4} (20,900 \text{ acfm}) (19 \text{ inches water})}{0.60} = 77.4 \text{ kW}$$

\*A fluid-bed catalytic incinerator is used because the waste gas contains a chlorinated compound which would poison the catalyst in a fixed-bed incinerator.

\*\*Computed from inlet waste-gas flow rate (20,000 scfm) at preheater inlet temperature (100°F).

### Catalytic Incinerator

$$Power_{fan} = \frac{1.17 \times 10^4 (20,900 \text{ acfm}) (23 \text{ inches water})}{0.60}$$

$$= 93.7 \text{ kW}$$

The annual electricity costs would be the products of these usages, the annual operating hours, and the electricity cost (\$/kWh), or:

$$\text{Electricity Cost (Thermal)} = 77.4 \text{ kW} \times 8,000 \text{ hours/yr} \times \$0.059/\text{kWh}$$

$$= \$36,500 \text{ per yr}$$

$$\text{Electricity Cost (Catalytic)} = 93.7 \text{ kW} \times 8,000 \text{ hours/yr} \times \$0.059/\text{kWh}$$

$$= \$44,200 \text{ per yr}$$

The catalyst replacement costs and scheduling are highly variable and depend on the nature of the catalyst, the amount of "poisons" and particulates in the gas stream (including the auxiliary fuel), the temperature history of the catalyst, and the design of the unit. It is impossible to predict the costs in a general sense. However, noble metal monolith catalysts operating on pure hydrocarbon gases in air will last longer than fluid-bed base metal catalysts operating on chlorinated hydrocarbons in air. Noble metal catalysts are also more expensive than base metal oxide catalysts. The catalyst life for many field units is from 1 to 4 years. The cost, in April 1988 dollars, of the replacement catalyst must be obtained from the vendor, but it may be estimated at \$3,000/ft<sup>3</sup> for noble metal catalysts and \$650/ft<sup>3</sup> for base metal oxide catalysts. For the example case, the catalyst is a base metal oxide because the waste gas contains a chlorinated compound. We will assume a two year catalyst life. Knowing that the catalyst volume is 39 ft<sup>3</sup> (Section 3.4.3) and using a cost of \$650/ft<sup>3</sup> and a capital recovery factor of 0.5531 (2-year life at a 7% interest rate), the annual expense for catalyst replacement is

$$\text{Annual Catalyst Replacement Cost} = 39 \text{ ft}^3 \times 650 \frac{\$}{\text{ft}^3} \times 0.5531 \times 1.08$$

$$= \$15,100 \text{ per year}$$

(The " 1.08" factor covers the freight and sales tax for the replacement catalyst.)

To calculate the fuel or electricity annual cost, multiply the fuel usage rate (scfm) or the electricity usage rate (kW) by the total hours per year of operation (e.g., 333 d/yr x 24 h/d = 8,000 h/yr) and by the appropriate unit cost (e.g., \$/scfm for fuel and \$/kWh for electricity).

For the example cases, the fuel costs can be calculated from the fuel usage rates and the natural gas unit cost of \$0.00330/scf. For the thermal incinerator example, the annual fuel cost is calculated as follows:

$$\begin{aligned}\text{Annual Fuel} &= 0.00330 \frac{\$}{\text{scf}} \times 167 \frac{\text{scf}}{\text{min}} \times 60 \frac{\text{min}}{\text{hr}} \times 8,000 \frac{\text{hr}}{\text{yr}} \\ \text{Cost, Thermal} &= \$264,500 \text{ per year}\end{aligned}$$

For the catalytic incinerator example, the annual fuel cost is found similarly:

$$\text{Annual Fuel Cost, Catalytic} = \$63,400 \text{ per year}$$

Operating and maintenance labor are estimated as 0.5 hours per 8-hour shift each, supervisory labor at 15% of operating labor, and maintenance material as 100% of maintenance labor.

### 3.5.2.2 Indirect Annual Costs

The indirect (fixed) annual costs include capital recovery, overhead, and property taxes, insurance, and administrative (G&A) charges. The last three of these can be estimated at 1%, 1%, and 2% of the total capital investment, respectively. The system capital recovery cost is based on an estimated 10-year equipment life. (See Section 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) The system capital recovery cost is the product of the system capital recovery factor (CRF) and the total capital investment (TCI) less the purchased cost of the catalyst ( $C_{cat} \times 1.08$  where the 1.08 is for freight and sales tax). These values calculated for the example cases are given in Table 3.10.

### 3.5.3 Cost Comparison for Example Case

The example VOC stream defined in Section 3.4.1 serves to illustrate some typical characteristics of thermal and catalytic systems. The total annual costs shown in Table 3.10 show that the catalytic system's auxiliary fuel costs are significantly lower than those of the thermal unit. The disparity is enough to offset the higher capital costs of the catalytic incinerator over the assumed 10-year lifetime of the units. Two factors that should be noted in the comparison of these two systems are (1) the 98 percent level of destruction met by the thermal incinerator may be difficult to reach by the catalytic system (this may be important in some cases), and (2) the example waste stream is of particularly low heating value (4 Btu/scf) which favors the catalytic system due to the lower auxiliary fuel requirements.

## 3.6 Acknowledgements

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## **Appendix 3A**

### **Properties of Selected Compounds**



**Table 3.12.** Limits of Flammability of Combustible Organic Compounds in Air at Atmospheric Pressure, Room Temperature\*

<b>Compound</b>	<b>Molecular Weight</b>	<b>LEL<sup>a</sup>, vol. %</b>	<b>UEL<sup>b</sup>, vol. %</b>
Methane	16.04	5.00	15.00
Ethane	30.07	3.00	12.50
Propane	44.09	2.12	9.35
Butane	58.12	1.86	8.41
Pentane	72.15	1.40	7.80
Hexane	86.17	1.18	7.40
Octane	114.23	0.95	
Nonane	128.25	0.83	
Decane	142.28	0.77	
Ethylene	28.05	2.75	28.60
Propylene	42.08	2.00	11.10
Acetylene	26.04	2.50	80.00
Cyclohexane	84.16	1.26	7.75
Benzene	78.11	1.40	7.10
Toluene	92.13	1.27	6.75

\*Reference [14]

<sup>a</sup>Lower Explosive Limit

<sup>b</sup>Upper Explosive Limit

**Table 3.13:** Molar Heat Capabilities of Gases at Zero Pressure\*

$$C_p = a + bT + cT^2 + dT^3; T \text{ in } ^\circ\text{K}$$

$$C_{pm} = \frac{\int_{T_1}^{T_2} C_p dT}{(T_2 - T_1)}$$

$$C_p \text{ in calories/g moles } ^\circ\text{K} \quad \text{Btu/lb mole } ^\circ\text{R}$$

Compound	a	b x 10 <sup>2</sup>	c x 10 <sup>6</sup>	d x 10 <sup>9</sup>	Temperature Range, °K
Methane	4.750	1.200	0.3030	-2.630	273-1500
Ethane	1.648	4.124	-1.530	1.740	273-1500
Propane	-0.966	7.279	-3.755	7.580	273-1500
Butane	0.945	8.873	-4.380	8.360	273-1500
Pentane	1.618	10.85	-5.365	10.10	273-1500
Hexane	1.657	13.19	-6.844	13.78	273-1500
Cyclopentane	-12.957	13.087	-7.447	16.41	273-1500
Cyclohexane	-15.935	16.454	-9.203	19.27	273-1500
Benzene	-8.650	11.578	-7.540	18.54	273-1500
Toluene	-8.213	13.357	-8.230	19.20	273-1500
Nitrogen	6.903	-0.03753	0.1930	-0.6861	273-1800
Oxygen	6.085	0.3631	-0.1709	0.3133	273-1800
Air	6.713	0.04697	0.1147	-0.4696	273-1800
Carbon dioxide	5.316	1.4285	-0.8362	1.784	273-1800

**W**

\* Reference [26]



## **Appendix 3B**

### **Design Procedure for Non-Recuperative Thermal Incinerators**

Not all thermal incinerators are equipped with recuperative heat exchangers to transfer energy from the flue gas stream to the incoming waste gas stream. These non-recuperative units use other mechanisms to recovery flue gas energy. One of these types is the *regenerative* incinerator. As discussed in Section 3.2.1.3, a regenerative incinerator accomplishes energy recovery by conveying the flue gas through a ceramic bed which captures a portion of the stream's enthalpy. After a switching mechanism is engaged, the incoming waste gas passes through this hot bed and is warmed to its ignition temperature. This process is illustrated in Figure 3.2.

While we can determine the stream inlet and outlet temperatures for a recuperative heat exchanger fairly accurately, we cannot always do so for a regenerative incinerator bed. For one thing, these beds do not behave like typical heat exchangers. The bed temperature profiles are often difficult to predict. More importantly, because regenerative incinerators do *not* operate at steady state conditions, the temperatures within the beds and many other parts of the unit vary with time. For that reason, it is more convenient to view the entire regenerative incinerator as a "black box" into which waste gas and auxiliary fuel flow and from which flue gas emanates. Around this black box we may make mass and energy balances. In this way, we need not make any assumptions about what occurs *inside* the incinerator regarding temperatures, flowrates, or other stream parameters.

However, to determine the auxiliary fuel requirement for regenerative incinerators via the procedure shown in this appendix we have to make two key assumptions, viz.: (1) the temperatures and flowrates of all streams entering and leaving the incinerator are at steady state and (2) the combustion temperature (and by inference, the heat loss fraction) are constant as well. The other assumptions will be addressed in the following design steps:

**Steps 1 to 4:** These are the same as those for thermal recuperative and catalytic incinerators. (See Section 3.4.1.)

**Step 5t - Establish the incinerator operating temperature:** Because their designs are more resistant to thermal stresses and because they can achieve very high heat recoveries, regenerative incinerators are usually operated at higher temperatures than recuperative units. Consequently, higher VOC destruction efficiencies are achieved. Operating temperatures of 1800 to 2000 °F are typical.

**Step 6t - Calculate the waste gas temperature at the exit of the preheater:** As explained above, regenerative incinerators do not employ preheaters. The preheating is done by and within the ceramic beds. Moreover, because the mass and energy balances are made around the entire unit, we do not need to know the temperature of the preheated waste gas to calculate the auxiliary fuel requirement.

**Step 7t - Calculate the auxiliary fuel requirement,  $Q_{af}$ :** Because a regenerative incinerator recovers nearly all of the energy from the combustion (flue) gas, its auxiliary fuel requirement is usually lower than that for a recuperative incinerator. However, as discussed above, this fuel requirement is determined via mass and energy balances taken around the entire unit, not just the combustion chamber. Consider the following diagram:

$$\begin{aligned}
 & \text{Flue gas} \\
 & T \\
 & * (fo) \\
 & * \\
 & R \\
 & +)))))))))))))))))))))) , \\
 & * \\
 & * \text{ Incinerator unit} \\
 \text{Aux. fuel S))))Q} & * (\text{any type}) * S)))) \text{ Waste} \\
 & * \\
 & * \text{ gas} \\
 & * (\text{wi}) \\
 & .)))))))))))))))))))))) -
 \end{aligned}$$

Taking mass and energy balances around the incinerator, we obtain:

### Mass balance:

$$\text{Mass in} = \text{Mass out}$$

$$\text{Mass fuel} + \text{Mass waste gas} = \text{Mass flue gas}$$

$$\rho_{af} Q_{af} + \rho_{wi} Q_{wi} = \rho_{fo} Q_{fo} \quad (3B-1)$$

### Energy balance:

Next, we take an energy balance around the incinerator unit:

$$\text{Energy in} - \text{Energy out} + \text{Energy generated} = 0$$

The terms of the energy balance equation are the inlet waste gas and outlet flue gas enthalpies ( $H_{wi}$  and  $H_{fo}$ , respectively), the energy loss ( $H_L$ ), and the waste gas VOC and fuel (natural gas) heat contents ( $H_{cwi}$  and  $H_{caf}$ , in turn):

$$H_{wi} - (H_{fo} + H_L) + (H_{cwi} + H_{caf}) = 0 \quad (3B-2)$$

The variables comprising each of the terms in this energy balance equation are listed in Table 3.6. They are:

$$H_{wi} = \rho_{wi} Q_{wi} C_{pmwi} (T_{wi} - T_{ref})$$

$$H_{fo} = \rho_{fo} Q_{fo} C_{pmfo} (T_{fo} - T_{ref})$$

$$H_L = \eta \rho_{fi} Q_{fi} C_{pmfi} (T_{fi} - T_{ref})$$

$$H_{cwi} = \rho_{wi} Q_{wi} (-\Delta h_{cwi})$$

$$H_{caf} = \rho_{af} Q_{af} (-\Delta h_{caf})$$

where:  $\eta$  = energy loss from combustion chamber (fractional)  
 $T_{fi}$  = combustion temperature (°F)

We next substitute these variables into eq. (3B-2) and solve for the fuel mass rate ( $\rho_{af}Q_{af}$ ). When doing so, we make the following assumptions:

- The streams flowing to and from the incinerator are at **steady state conditions**.
- The auxiliary air requirements are zero.
- The ambient, reference, and fuel inlet temperatures are equal (77 °F). (This assumption results in the inlet fuel stream having a zero enthalpy.)
- The heat capacities of the gas streams to and from the unit are approximately the same, regardless of *composition*.
- The *mean* heat capacities of the streams above the reference temperature (77 °F) are approximately equal, regardless of *temperature*. Further, the mean heat capacity of the waste gas/flue gas stream entering/leaving the incinerator is evaluated at the average of the inlet ( $T_{wi}$ ) and combustion ( $T_{fi}$ ) temperatures. That is,  $C_{pmwi} = C_{pmfi} = C_{pmfo} = C_{pm}$ .

When we do all this, we get the following expression:

$$\begin{aligned} & \rho_{wi}Q_{wi}C_{pm}(T_{wi} - T_{ref}) - [\rho_{fo}Q_{fo}C_{pm}(T_{fo} - T_{ref}) + \eta\rho_{fi}Q_{fi}C_{pm}(T_{fi} - T_{ref})] + \\ & \quad \text{(Energy in)} \qquad \qquad \qquad \text{(Energy out)} \\ & [\rho_{wi}Q_{wi}(-\Delta h_{cwi}) + \rho_{af}Q_{af}(-\Delta h_{caf})] = 0 \\ & \quad \text{(Energy generated)} \end{aligned}$$

Substitution for  $\rho_{fo}Q_{fo}$  per eq. 3B-1 above yields:

$$\begin{aligned} & \{\rho_{wi}Q_{wi}C_{pm}(T_{wi}-T_{ref})\} - \{\eta C_{pm}(\rho_{af}Q_{af} + \rho_{wi}Q_{wi})(T_{fi}-T_{ref}) + \\ & C_{pm}(\rho_{af}Q_{af} + \rho_{wi}Q_{wi})(T_{fo}-T_{ref})\} + \{\rho_{wi}Q_{wi}(-\Delta h_{cwi}) + \rho_{af}Q_{af}(-\Delta h_{caf})\} = 0 \end{aligned}$$

Finally, solving for  $\rho_{af}Q_{af}$ , the auxiliary fuel mass rate (lb/min):

$$\rho_{af}Q_{af} = \frac{\rho_{wi}Q_{wi}\{C_{pm}[\eta(T_{fi} - T_{ref}) + (T_{fo} - T_{wi})] - (-\Delta h_{cwi})\}}{\{(-\Delta h_{caf}) - C_{pm}[\eta(T_{fi} - T_{ref}) + (T_{fo} - T_{ref})]\}} \quad (3B-3)$$

Equation (3B-3) provides the auxiliary fuel requirement for any type of thermal incinerator, as it is independent of any intermediate variables, such as the temperature of the preheated waste gas. Clearly, this equation can be used with regenerative incinerators, as long as the above-stated assumptions hold.

The heat loss fraction ( $\eta$ ) will vary according to the incinerator type, how the incinerator components are configured in the unit, the construction materials, the type and amount of insulation, and other factors. For instance, for recuperative incinerators,  $\eta$  is approximately 0.10. The  $\eta$  for regenerative incinerators is considerably lower, however. There are two reasons for this. First, the components of a regenerative incinerator—combustion chamber, ceramic beds, etc.—are housed in a single enclosure, while in a recuperative incinerator the combustion chamber, heat exchanger, and interconnecting ductwork are housed separately, thus offering more heat transfer area. Second, because regenerative units are lined with ceramic, they are better insulated than recuperative incinerators.

To gain an estimate of this heat loss fraction, we contacted two regenerative incinerator vendors. [27,28] Based on the heat loss data that they supplied, we calculated  $\eta$  values ranging from **0.002 to 0.015 (0.2 to 1.5%)**. These values varied according to the incinerator configuration (vertical or horizontal), the waste gas flow rate, the ambient temperature, and the wind speed.

**Step 8t - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame:** As explained in Section 3.4.2, only a small amount ( $< 5\%$  of the total energy input) is needed to stabilize the burner flame. With recuperative incinerators, the auxiliary fuel requirement is usually much larger than the burner stabilization requirement, so that this constraint rarely comes into play. With regenerative incinerators, however, the auxiliary fuel requirement may be as low or lower than the fuel needed to stabilize the burner. Therefore, it is important to compare these two requirements. This comparison is made via equations 3.19 and 3.20. If the auxiliary fuel is less, the minimum fuel requirement would be set at 5% of the total energy input.

**Step 9t - Calculate the flue gas volumetric flow rate,  $Q_{fi}$ :** As with thermal recuperative incinerators, the regenerative incinerator flue gas flow rate is the rate used to size and cost the unit. Measured at standard conditions (1 atmosphere and 77° F),  $Q_{fi}$  is the sum of the inlet waste gas ( $Q_{wi}$ ) and fuel ( $Q_{af}$ ) flow rates. But since  $Q_{af}$  for regenerative units is small compared to  $Q_{wi}$ , the waste gas and flue gas flows should be virtually identical.



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